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Catalytic Reduction of Nitrogen Oxides with Ammonia: Utility Pilot Plant Operation

Environics, Inc, Newport Beach, Calif

Prepared for

Industrial Environmental Research Lab, Research Triangle Park, N C

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16. ABSTRACT: The report describes work to demonstrate, on a utility pilot plant scale, the performance, reliability, and practicability of reducing nitrogen oxides (NOx) emissions from steam boilers by reduction of NOx with ammonia over a platinum catalyst. A utility pilot plant treating a slipstream from an operating electric utility boiler, equivalent to approximately 1.5 MW output, was designed, installed, and tested on gas and oil fuel. Laboratory pilot plant testing supplemented the utility pilot plant testing. Results of gas-firing the utility pilot plant, together with laboratory pilot plant test results, indicated that the catalytic reduction system has consistently provided 85-90% NOx removal for over 4000 hours, with no significant performance loss at a 50,000 per hour space velocity. Results of oil-firing the utility pilot plant, together with laboratory pilot plant tests results, indicated that this system could provide at least 65% NOx removal when the flue gas temperature is above 575 F. Preliminary cost estimates indicate a capital expenditure of less than $11 per kW plant capacity, with an operating expenditure of less than 0.02¢ per kWhr for a full-scale system, based on 1974 dollars and a gas-fired 480 MW plant.

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October 1976

CATALYTIC REDUCTION OF NITROGEN OXIDES WITH AMMONIA: UTILITY PILOT PLANT OPERATION

by

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Program Element No. EHE624

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We wish to acknowledge the contribution of the Los Angeles Department of Water and Power (LADWP), who allowed the installation and operation of the pilot plant at their Valley Steam Plant, Unit 4, despite the attendant inconvenience. We wish to thank LADWP management and particularly Messrs. Hans Sonderling, Wesley Pepper, Charles Sun-Woo, Irvine Tuttle and Lloyd Reeve for their cooperation and assistance.

We also wish to acknowledge the significant contributions of the Environmental Protection Agency, particularly Messrs. Richard D. Stern, Roger Christman, Jim Wingo, Luis Garcia, Kenneth Baker, and Dale Denny.

Some of the small scale work performed outside the scope of this program, reported in Section 3.3 as background information, was performed with the support of the Southern California Edison Company and the San Diego Gas and Electric Company.

SECTION 1

SUMMARY

1.1 The contract #68-02-0292 was effective, beginning January 1973, although work actually began in October 1972.

1.2 PROGRAM OBJECTIVES

The principal objective of the contract was to treat the flue gas of a gas fired boiler to demonstrate, on a utility pilot plant scale, the performance and practical aspects of a NOx reduction system, using a platinum catalyst with ammonia as the reductant. Because of the decreasing availability of natural gas for utility boilers, the program was expanded to include tests on oil firing.

More specifically, the primary objective of this program was to demonstrate that the combined emissions of NOx and NH_3 could be controlled to less than 50 ppm.

The second objective was to establish the useful life of the platinum catalyst. Chemical activity of the catalyst could conceivably have been reduced by poisoning, or its physical stability could have been impaired by exposure to the flue gas environment. A projected catalyst life of 12 months or more was desired.

The third objective was to establish optimum temperatures, flow rates of NH_3, size of required catalyst bed, and system control parameters, for future use in designing full-scale flue gas treatment to remove NOx.

The fourth objective was to determine the economic feasibility of utilizing this process to reduce the NOx emissions from a 500 MW gas-fired utility boiler to below 50 ppm. Capital investment cost estimates, including the installation and/or

modification of boiler components, and the initial cost and replacement cost of the catalyst, and operating cost estimates, including maintenance, as well as cost of the NH_3 used as a reductant, were desired.

1.3 THE UTILITY PILOT PLANT

Negotiations between Environics and the Los Angeles Department of Water and Power began prior to the effective date of the contract. An agreement was reached with the Los Angeles Department of Water and Power (LADWP) in October 1972.

In this agreement, Boiler #4, having 170 MW capability, at the San Fernando Facility of the LADWP, was designated as the boiler for the pilot plant. Additionally, LADWP agreed to furnish engineering services and approve modifications to the boiler that may be necessary to accommodate the pilot plant. LADWP was to receive all progress reports and the final report. Periodic reviews were conducted with LADWP engineering personnel and operating personnel at the San Fernando facility. The pilot plant was sized to handle a slipstream equivalent to 1.5 MW output (150,000 SCFH). Its design was based upon Environics' previous small scale test work.

1.3.1 Catalyst Selection

Environics' previous small scale tests had indicated that a 0.3% platinum catalyst, supported on a high surface area washcoated ceramic honeycomb, would provide high activity for NOx removal in steam boiler flue gas with a relatively low pressure drop through the catalyst bed. Thus, selection of the pilot plant catalyst was limited to such a catalyst, with the objective of selecting the proper size support, to provide acceptable levels of activity and pressure drop for a utility boiler. Testing was performed to determine pressure drop and activity characteristics of several catalyst/support configurations.

A realistic space velocity of 50,000 v/v/hr is targeted. The catalyst volume is therefore 7 cu. ft.

The unique feature of the utility pilot plant is the incorporation of the catalyst in the air preheater by replacing the heat transfer elements inside the air preheater with the catalyst. This feature makes the use of catalytic reduction to treat the flue gas simple and practical, particularly in retrofitting an existing power plant. Because of the closely coupled arrangement of the power plant components, there is normally no room for the addition of a separate catalytic converter. For a new plant, a separate catalytic converter to treat its flue gas is feasible.

1.4 GAS-FIRED TESTING

1.4.1 The gas-fired testing began with small scale-long duration testing. The small scale test boiler ran for 175 days (4200 hours) with no sign of catalyst deterioration. Although there were several instances of soot accumulating on the catalyst, these were considered due to upsets in the operation of the small boiler, not to its normal function, and the catalyst was easily regenerated. At the end of the test period, 90% NOx removal was again obtained, with an ($NO_x + NH_3$) exit concentration less than 50 ppm. The analysis for trace products found no detectable contaminants from side reactions. The small scale-long duration testing is discussed in detail in Section 6.

1.4.2 The results of the utility pilot plant operation on gas firing were equally encouraging. The pilot plant was operated on gas firing for over 2000 hours and the catalyst showed no sign of deterioration. Although steady state conditions were difficult to maintain, it appeared that the pilot

plant catalyst maintained 85-90% NOx removal under optimum conditions of 490°F and 200 ppm NH_3.

That the pilot plant could not accumulate more running time during the gas fired test period was due mainly to problems with the boiler. These problems included long periods when gas was not available, lengthy major maintenance shutdowns, and minor maintenance shutdowns to repair tube leaks and other malfunctions. The pilot plant itself, because of its relative simplicity, was operated for essentially the maximum time possible during the test period. The details are discussed in Section 8.

A summary of the optimum operation conditions and its performance are given below:

Catalyst	Pt, 0.03% by weight.
Ammonia inlet	200 ppm.
Ammonia exit	Less than 15 ppm.
Temperature	490°F.
Pilot plant flow rate	150,000 SCFH approximately.
Pilot plant power (equivalent)	1.5 MW
Space velocity	50,000 v/v/hr.
NOx removal efficiency	85-90%
Running time, accumulative	2000 hrs.

1.5 OIL-FIRED TESTING

1.5.1 Results of the oil fired testing, with the utility pilot plant and laboratory pilot plant, were not as encouraging as those on gas firing. Indeed, these results were somewhat contradictory to the results of Environics' previous small scale work.

The utility pilot plant was operated on oil firing for a total time of over 400 hours. During this period, the maximum NOx removal efficiency that was achieved was 65%; the average NOx removal efficiency was 50%. Because of the difficulties encountered in maintaining constant temperature in the pilot plant, it was not possible to conduct accurate parametric tests. However, it did appear that to obtain maximum NOx removal efficiency, a catalyst temperature between 550 and 600°F was required. In contrast to this, the previous small-scale data indicated that peak NOx removal efficiency of 80-95% could be obtained between 520 and 575°F. It is possible that higher NOx removal efficiency could have been obtained in the utility pilot plant if it would have been possible to maintain a constant, higher temperature. This is because of the relatively long time required to obtain steady state conditions in the presence of SO_2 (presumably due to adsorbtion/desorbtion of some NH_3/SO_2 compound on the catalyst surface). It is also possible, however, that the relatively poor results were due to the effect of species in the flue gas other than SO_2.

The other problems encountered during oil fired operation of the pilot plant were catalyst plugging with soot and ammonium sulfate accumulation. The plugging, observed on the leading edge of the catalyst, could probably be minimized by use of a larger cell size honeycomb support, countercurrent operation of the air preheater, and/or by periodic cleaning operations using compressed air or water. The white particulate accumulation could probably be minimized by use of an automatic NH_3 concentration control loop, to maintain an appropriate inlet NH_3 concentration as power plant load changes. The details are given in Section 9.

1.5.2 However, the laboratory pilot plant tests did show that the oil "poisoning," if any, is reversible; i.e., when

the boiler was switched to gas fuel, the catalyst gradually recovered its gas-fired performance.

1.6 COST ANALYSIS (GAS-FIRED BOILER)

Based on preliminary cost estimates that were performed, it was expected that capital cost for the full scale system would be below $15 per KW, and that operating costs would be below 0.1 cents per KW hour. Capital cost is quite sensitive to catalyst life, and thus it is expected that this program would enable much more accurate cost estimates than have heretofore been possible.

Based upon program results, the capital cost for a full scale catalytic reduction system is estimated to be $11 per KW plant capacity. The total operating cost associated with such a system is estimated to be 0.02¢ per kw-hr. These estimates were based on 1974 dollars and a gas fired 480 MW plant.

SECTION 2

CONCLUSIONS

The following conclusions can be made about the results of the pilot plant program.

1. Catalytic reduction is a practical method of reducing NOx emissions from a gas fired (natural or synthetic gas) steam boiler. Up to 90% reduction can be achieved, without emission of excessive unreacted NH_3. The catalyst can be conveniently placed in the air preheater wheel. Thus, the retrofitting of an operating utility steam boiler with a catalytic converter by replacing the heat transfer buckets with catalytic elements without any change of the external configuration is made simple, practical and inexpensive.

2. The effluent ammonia has consistently been less than 10 to 15 ppm.

3. The catalytic converter when using gaseous fuel consistently yielded 85-90% conversion at a space velocity of 50,000 v/v/hr.

4. Reliability of the catalytic converter system, using gaseous fuel, has been demonstrated for more than 4,000 hours of accumulative testing.

5. The controls associated with the catalytic converter system can be simple, using a feedback system, which senses either NOx or ammonia effluent or summation of both to regulate ammonia injection.

6. Catalytic reduction does not provide as great a degree of NOx control in an oil-fired plant as in a gas-fired plant. Perhaps 65 to 70% control may be achieved with oil. Installation of the catalyst in the air preheater provides a

desirable continuous regeneration effect. The soot accumulation problem can be solved by a soot remover located in the air preheater. The optimum NOx conversion temperature for oil fired boilers is above 575°F.

7. In a plant fired with gas and oil, alternately, high performance will return when firing is switched back to gas, from oil.

8. The capital cost for a full scale catalytic reduction system is estimated to be less than $11 per KW plant capacity. The total operating cost associated with such a system is estimated to be less than 0.02¢ per KW-hr. These estimates were based on 1974 dollars and a gas fired 480 MW plant.

SECTION 3

RECOMMENDATIONS

The following recommendations are made for continued development of a catalytic NOx reduction system leading to a large scale prototype pilot plant.

1. Further research should be performed, on a laboratory pilot plant scale, to optimize the system for oil firing. Such research should include tests of platinum and transition metal catalysts, with particular attention to the beneficial effects of operation in a rotating air preheater wheel. A practical system must be capable of operating on both oil and gas fuels.

2. Simultaneous with the laboratory pilot plant research, the utility pilot plant testing should be continued. If possible, negotiation with LADWP to relocate the present pilot plant either at the Valley Steam Plant or other location so that the utility pilot plant can be operated at higher temperatures.

3. A full scale prototype catalytic converter system should be initiated for gas-fired plants such as LADWP Scattergood #3 or other LADWP's smaller plants. This recommendation is directly applicable for synthetic gas plants of the future.

4. Long-term need is in the coal-fired plants. Research should begin in catalytic removal of oxides of nitrogen, associated with coal-firing.

SECTION 4

INTRODUCTION

4.1 NITROGEN OXIDE EMISSION AND CONTROL

Furnaces, stationary engines and turbines burning fossil fuels emit large quantities of nitrogen oxides. The sources of these emissions are power plants, oil refineries, gas pumping stations, and oil and gas producing facilities.

Combustion modification techniques appear to be the quickest and least expensive methods to accomplish a first-order reduction of NOx emissions from most stationary sources. The techniques which have been used include low excess air combustion, staged combustion, flue gas recirculation, steam or water injection, and various combinations of these techniques. A realistic control level using such techniques is 100 ppm NOx, for gas-fired furnaces, and higher levels for staionary engines which burn natural gas, gasoline, jet fuel, fuel oil, or coal.

However, the control of NOx emissions by combustion modification may not be applicable to those special processes which require very high temperatures (i.e., glass manufacture), compact furnaces or processes burning fuels containing large amounts of organically bound nitrogen. Furthermore, it may be desirable, at least in metropolitan areas, to reduce NOx emissions more than is possible with combustion modification.

One method which has the potential for significant reduction of NOx emissions from stationary sources is catalytic reduction of NOx with ammonia. In this method, a controlled amount of NH_3 is mixed with the boiler flue gas, and the mixture is passed over a catalyst at the proper conditions of flowrate and temperature. The catalyst promotes the following

reactions which convert the NOx and NH_3 to harmless nitrogen and water vapor:

$$6\ NO + 4\ NH_3 \longrightarrow 5N_2 + 6H_2O \qquad (1)$$

$$6\ NO_2 + 8\ NH_3 \longrightarrow 7\ N_2 + 12\ H_2O \qquad (2)$$

Ammonia is considered a selective reductant for NOx since it reacts via Reactions (1) and (2) in preference to the reactions with oxygen given below:

$$3\ O_2 + 4NH_3 \longrightarrow 2\ N_2 + 6\ H_2O \qquad (3)$$

$$5\ O_2 + 4NH_3 \longrightarrow 4\ NO + 6\ H_2O \qquad (4)$$

(In practice, Reactions (3) and (4) also proceed, but to a lesser extent, or at a higher temperature, than Reactions (1) and (2).)

Catalytic reduction of NOx with NH_3 can be employed as an alternative to, or in combination with, combustion modification, and can reduce NOx emissions by as much as 95%.

4.2 OBJECTIVE OF PILOT PLANT PROGRAM

The objective of the work reported herein was to demonstrate, on a utility pilot plant scale, the practicability of catalytic reduction of NOx with NH_3, using a platinum catalyst. The utility pilot plant was sized to handle a slipstream from a utility steam boiler, equivalent to approximately 1.5 MW output ($\sim$ 150,000 SCFH). Its design was based upon Environics' extensive previous small scale test work.

The pilot plant was installed at the Los Angeles Department of Water and Power Valley Steam Plant, Unit No. 4, a 170 MW steam boiler, equipped for natural gas and fuel oil firing.

Initially, it was desired to limit operation of the pilot plant to gas firing only. Subsequently (in view of the decreasing availability of natural gas fuel for utility boilers), the

program was expanded to include tests on oil firing, as well.

More specifically, the primary objective of this program was to demonstrate that the combined emissions of NOx and NH_3 could be controlled to less than 50 ppm.

The second objective was to establish the useful life of the platinum catalyst. Chemical activity of the catalyst could conceivably have been reduced by poisoning, or its physical stability could have been impaired by exposure to the flue gas environment. A projected catalyst life of 12 months or more was desired.

The third objective was to establish and to confirm previously obtained operating parameters such as optimum temperatures, flow rates of NH_3, size of required catalyst bed, and system control parameters, for future use in designing full-scale flue gas NOx removal equipment. Space velocity goal was 50,000 v/v/hr.

The fourth objective was to determine the economic feasibility of utilizing this process to reduce the NOx emissions from a 500 MW gas fired utility boiler to below 50 ppm. Capital investment cost estimates, including the installation and/or modification of boiler components, and the initial cost and replacement cost of the catalyst, and operating cost estimates, including maintenance, as well as cost of the NH_3 used as a reductant, were desired.

Based on preliminary cost estimates that were performed, it was expected that the capital cost for a full scale system would be below $15 per KW, and that operating costs would be below 0.1 cents per KW hour. Capital cost is quite sensitive to catalyst life, and thus this program would enable much more accurate cost estimates than have heretofore been possible.

The contract #68-02-0292 was initiated during October 1972.

Accordingly, the work reported herein was begun on October 1, 1972. Prior to that time, however, Environics had conducted extensive testing of the catalytic reduction system, on a small scale (100 - 3000 SCFH). The small scale work undertaken before contract initiation is summarized as follows.

4.3 ENVIRONICS' PREVIOUS WORK

4.3.1 Gas Fired Small Scale Work

Prior to this program, Environics had conducted an extensive literature search, followed by laboratory research on catalytic reduction of NOx with NH_3 using numerous catalysts in several different sized small scale test units. Some of the catalyst tested are described in Table 4.1. Some small scale field tests on actual boilers and gas turbines were also performed. The results of this work are described below.

Initial laboratory work was performed using bottled gas mixtures of NO in nitrogen (with the addition of O_2 in some tests), and a one inch diameter converter. Flowrates with this bottled gas system were approximately 100-300 SCFH.

Results of these tests (summarized in Table 4.2) showed that 90 to 98% NOx removal could be achieved, using either platinum, copper oxides or vanadia based catalysts. This work demonstrated the feasibility of application of catalytic reduction with NH_3 to control NOx emission from steam boilers. Platinum catalysts were significantly more active than vanadia or copper oxide, giving 95% removal at approximately 40,000 v/v/hr space velocity compared to 95-98% removal at 10,000 v/v/hr for CuO, and 93% removal at 10,000 v/v/hr for V_2O_5.

The next stage of the small scale work was initiated in order to determine whether other components of flue gas (e.g., water vapor, CO and CO_2) would affect the behavior of the

various catalysts. In this stage of the program, a gas fired furnace was used to provide flue gas to a two inch diameter converter. Thus, the gas stream simulated the flue gas in a steam boiler.

The small scale test program concentrated on platinum catalysts, because of their high activity. Transition metal catalysts, such as copper oxide, manganese oxide, etc., were also tested extensively, however. Results of work with the platinum catalysts will be discussed herein.

Experiments were performed to determine the effects of temperature, space velocity and inlet NOx concentration, on NOx removal efficiency, with several platinum catalysts.

A typical result showing the effect of temperature on NOx removal efficiency is presented in Figure 4.1. As can be seen, peak NOx removal efficiency of as high as 98% was obtained in the temperature range of 400° - 500°F.

Results showing the effect of space velocity on NOx removal efficiency are shown in Figure 4.2.

Experiments were also performed to determine the effect of inlet NH_3 concentration on both NOx removal efficiency and NH_3 effluent concentration. Results are shown in Figure 4.3; for platinum catalyst, at 500°F.

The next phase of this small scale testing was to determine the effect of SO_2 on performance of the various catalysts. This work is described below.

4.3.2 SO_2 Experiments

It had been noted in the literature, and subsequently, that information on the effects of SO_2 on platinum (and other) catalysts was contradictory. In view of the increasing quantity of fuel oil, relative to natural gas, being used by electrical

generating stations, it was decided to direct the efforts of the small scale testing towards determining the effect of SO_2 on the NOx reduction system.

This work was performed using the test furnace system described above, by injecting SO_2 into the exhaust gas upstream of the converter.

Initial SO_2 tests were made using a platinum catalyst operating at other than optimum conditions (450 to 500 F, 50,000 hr^{-1} SV). The results were that, within several hours, the catalyst showed a marked decrease in performance, with NOx removal efficiency dropping to below 20%. It was eventually learned that this apparent SO_2 "poisoning" could be prevented by increasing the temperature of the catalyst above 575°F.

A typical result of some of these SO_2 experiments, with platinum catalysts, is presented in Figure 4.4, showing NOx removal efficiency as a function of catalyst age, with tests at various temperatures as indicated.

4.4 DISCUSSION

The purpose of the following comments is to put the small scale work described in Section 4.3, above, into the time perspective of the pilot plant work described in the remainder of this report.

At the time design of the pilot plant was started (October 1972), the literature search, bottled gas experiments, gas fired furnace experiments and a portion of the SO_2 experiments had been completed. The pilot plant was designed to operate only on natural gas fuel. Such operation was considered practical for a utility boiler in view of the fuel availability situation at that time.

The remainder of the SO_2 experiments was conducted during

the period in which the utility pilot plant was being designed and operated. In view of the results of that work, and the increasing shortfall in natural gas supply, the pilot plant was modified and operated on fuel oil firing, towards the end of the program described herein.

It should be noted that none of the small scale work described in Section 4.3 was performed as part of the utility pilot plant program.

The following sections discuss the design, operation, and results of the utility pilot plant program.

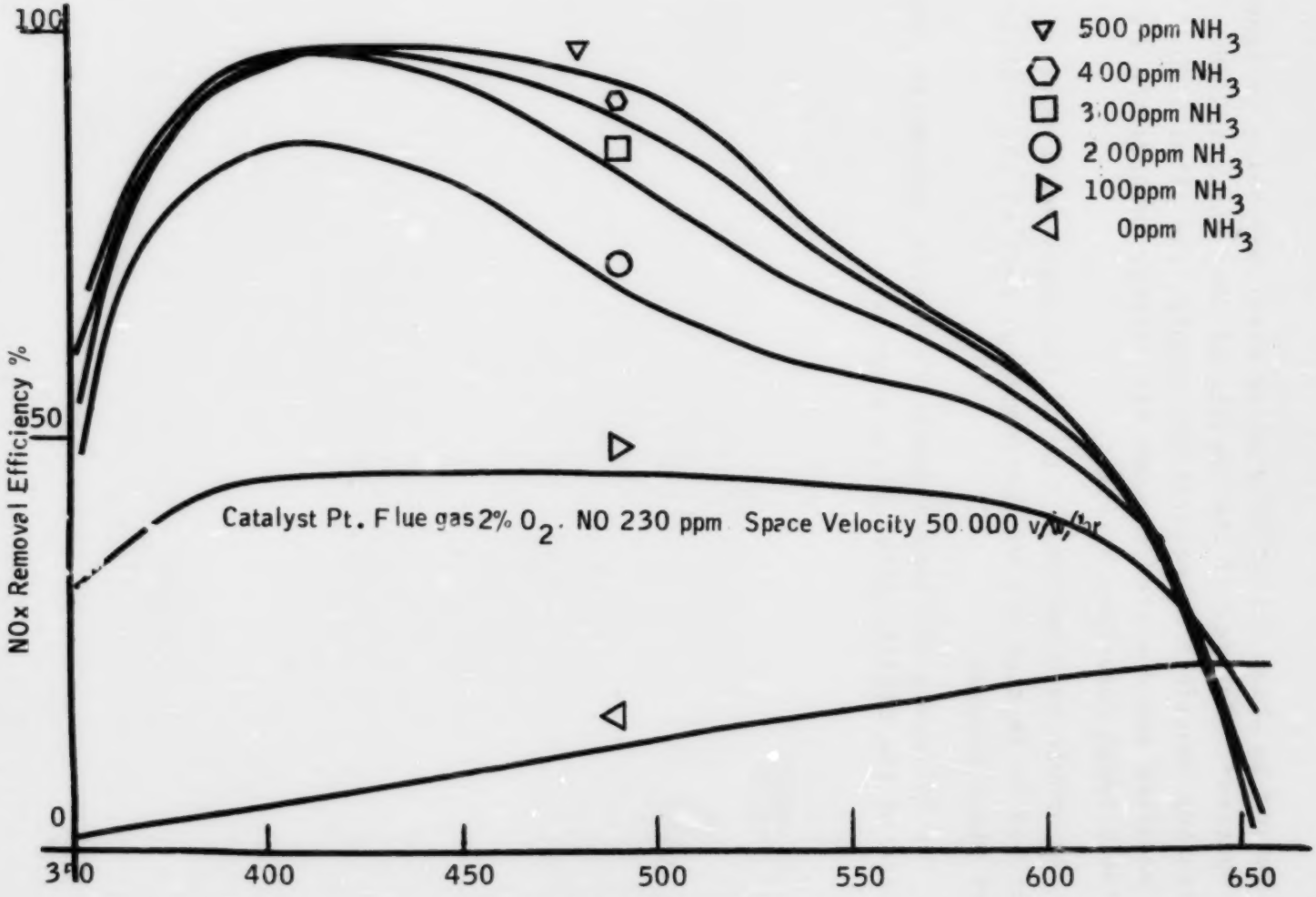


Figure 4.1 Temperature Vs NOx Removal Eff.

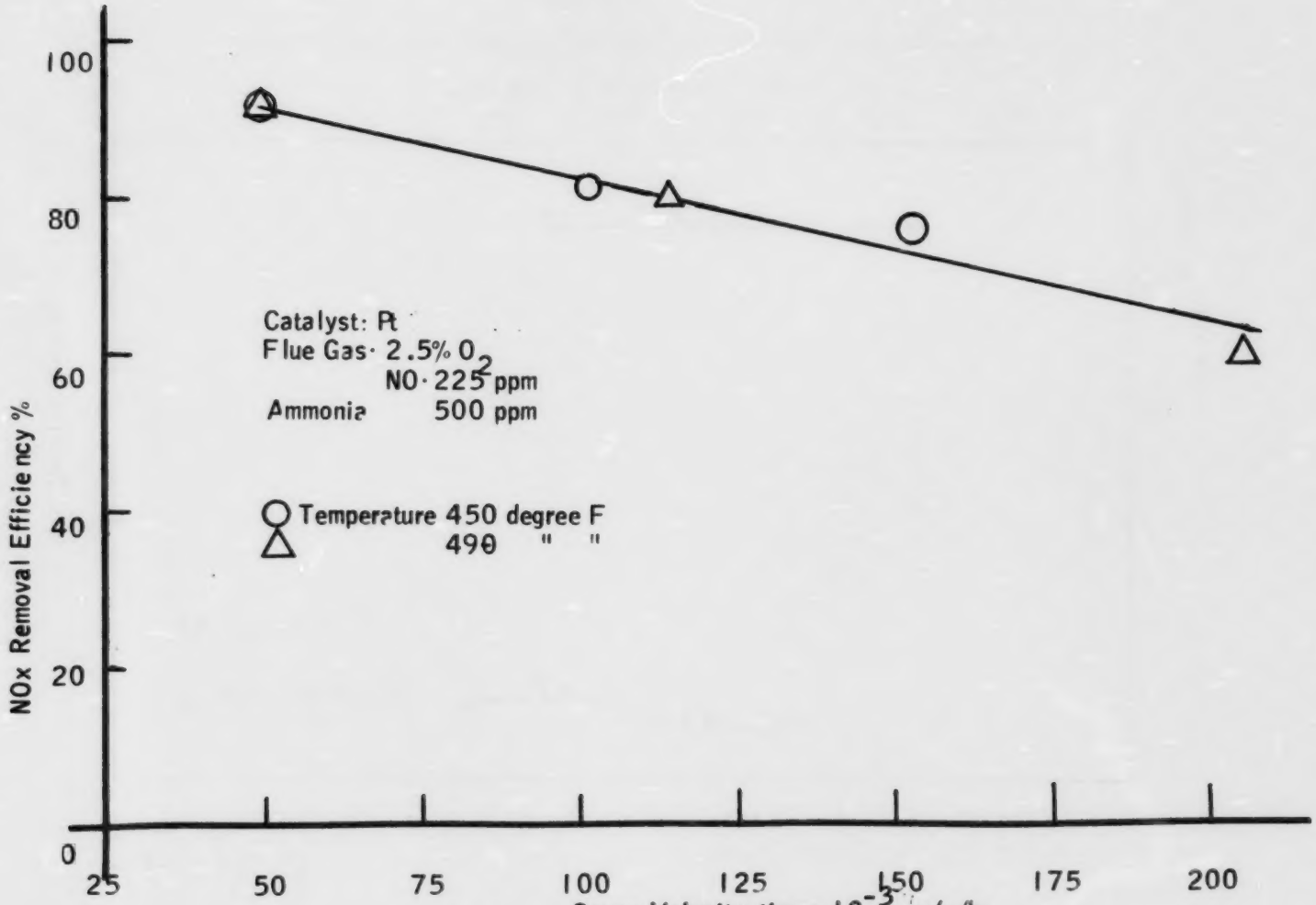
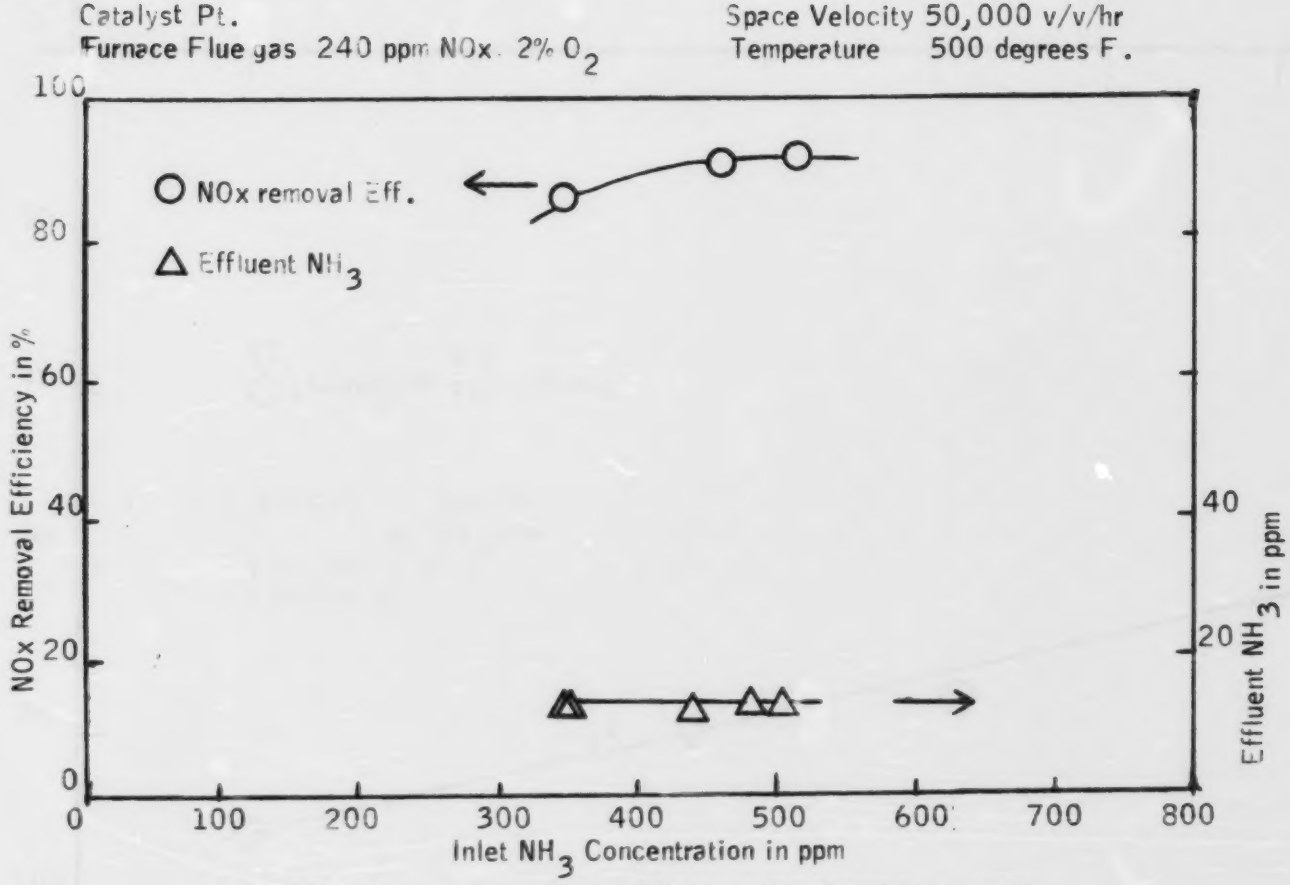
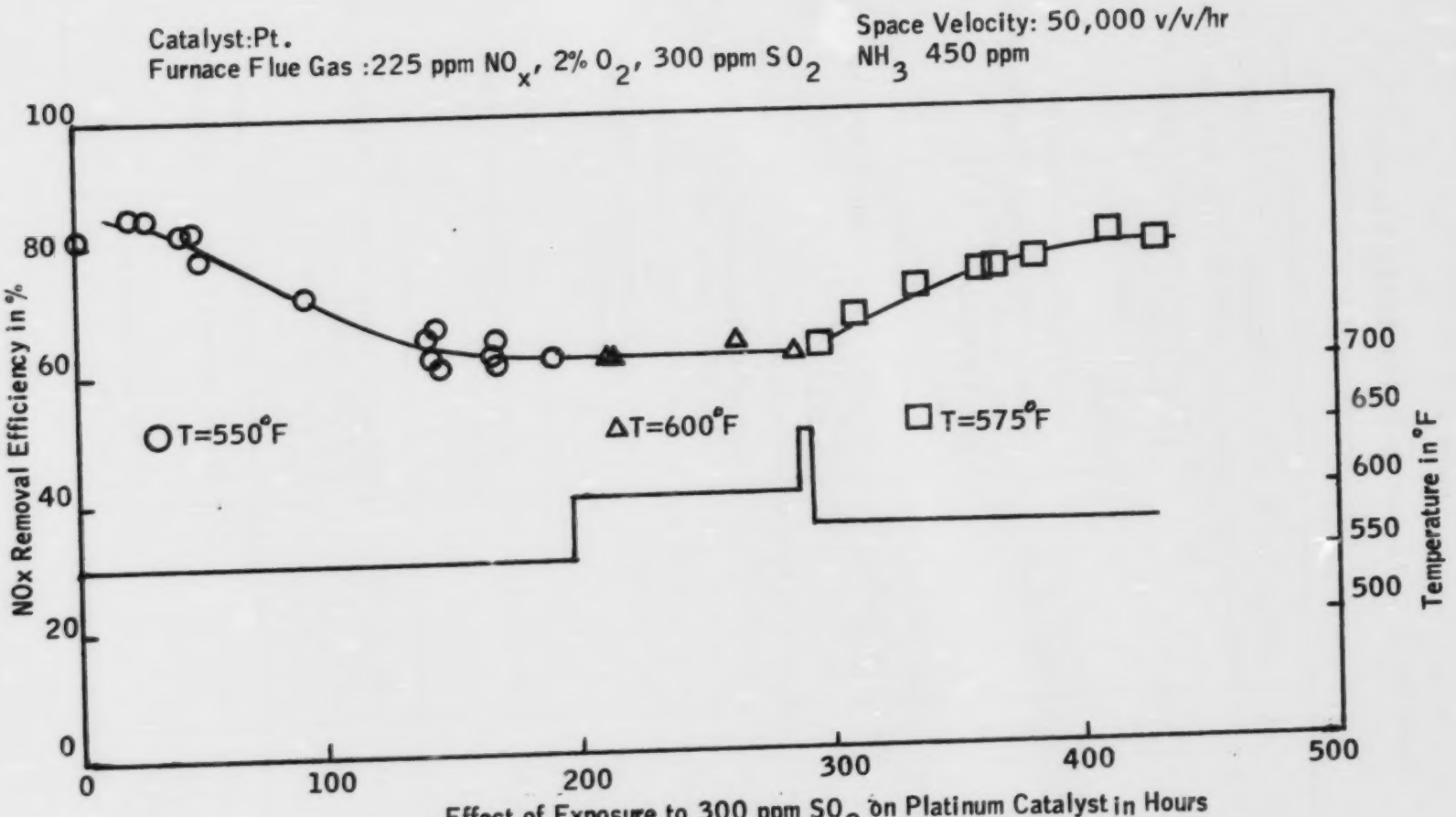


Figure 4.2 Removal Efficiency Vs Space Velocity



Removal Efficiency and Effluent NH_3 Vs Inlet NH_3 Concentration

Figure 4.3



Effect of Exposure to 300 ppm SO_2 on Platinum Catalyst in Hours

Figure 4.4

Table 4.1

Catalysts Included in Small Scale Testing

Designation	Composition	Support	Manufacturer
PtO2H - 1F2 - 1F4 - 1F5 - 1F6	Pt, 0.3%	Ceramic Honeycomb/ Al_2O_3 wash coat, 12 cpi* (American Lava)	Matthey Bishop
PtO2H - 1H1 - 1H3 - 1H4	Pt, 0.3%	Ceramic Honeycomb/ Al_2O_3 wash coat, 8 cpi (American Lava)	Matthey Bishop
PtO2H - 3A1	Pt + Ru, 0.3%	Ceramic Honeycomb, 1/8" cell (Du-Pont)	Matthey Bishop
PtO2H - 1G1	Pt, 0.3%	Ceramic Honeycomb, 1/8" cell (Hexcel)	Matthey Bishop
PtO3H - A1 - A2	Pt. 0.3%	Ceramic honeycomb, 12 cpi (American Lava)	Englehard
PtO6H - B1	Pt, 0.3%	Ceramic Honeycomb, 1/8" cell (Du-Pont)	Du Pont
PtO7H - A1	Pt, 0.3%	Ceramic Honeycomb, silica washcoat 8 cpi (American Lava)	Matthey Bishop
PtO4R - A1 - B1 - C1 - D1	Pt	Metal support D1-metal contains alumina	Matthey Bishop

Table 4.1 (continued)

Designation	Composition	Support	Manufacturer
Pt16G - A	Pt, 0.3%	Zeolite granules	Matthey Bishop
Pt20G - A	Pt, 0.3%	Active carbon granules	Matthey Bishop
Cu1OH	CuO/Cr_2O_3, 17%	Ceramic Honeycomb, 1/8" cell (DuPont)	DuPont
Cu11H	CuO	Ceramic Honeycomb, 1/8" cell (DuPont)	DuPont
Co15T Co16T	CoO	Tablets, 1/8" diameter, 1/8" long	Girdler
Cu09T Cu09T - 1	CuO 6.2% Co_2O_3 5% Aluminum hydroxide 33%	Tablets, 1/8" diameter 1/8" long	Harshaw
Cu08T	Barium promoted copper chromite ($Ba/CuO/Cr_2O_3$)	Tablets, 3/16" diameter 1/8" long	Girdler
Cu09H - A1	CuO, CoO	Ceramic Honeycomb, 1/8" cell (DuPont)	DuPont
La01 - P	Lanthanum oxide, Alumina	Tablets, 1/8" diameter, 1/8" long	Environics

Table 4.1 (continued)

Designation	Composition	Support	Manufacturer
M19H - A	MnO and others	Ceramic Honeycomb, 1/8" cell (DuPont)	DuPont
M17H - A	Monel, 100%	Pure Monel in honeycomb structure	Hexcel
V14E	V_2O_5 16% MoO_3 3%	Alumina extrusions, 3/16" diameter, 3/16" long	Harshaw

cpi = corrugations per inch

SUMMARY OF CATALYST PERFORMANCE: NO ($\sim$ 300 ppm) IN N_2

CATALYST	SPACE VELOCITY v/v/hr	TEMPERATURE RANGE (°F)	NOX REMOVAL EFFICIENCY
Cu07T (CuO/Cr_2O_3)	10,500(1)	600-700	95-98%
Cu08T ($Ba/CuO/Cr_2O_c$)	10,500(1)	600-700	95-98%
Pt01T (Pt)	17,500(2)	600-700	15-52%
Cu09T (CuO/CoO)	10,500(1)	600	99%
V14T (V_2O_5/MoO_3)	10,500(1)	600-650	75-93%
Pt02H (Pt)	40,000(3)	500	90%

(1) Catalyst bed dimensions = 1" dia. x 2.0" deep (pellets)

(2) Catalyst bed dimensions = 1" dia. x 1.2" deep (pellets)

(3) Catalyst bed dimensions = 3/4" dia. x 2.0" deep (honeycomb)

Table 4.2

SECTION 5

DESIGN OF UTILITY PILOT PLANT

Design of the utility pilot plant was conducted in several areas. These were (1) system design, (2) control system design, (3) ammonia injection system design, and (4) analytical techniques.

5.1 SYSTEM DESIGN

System design included overall design of the ductwork, selection of the catalyst support from the standpoint of activity and pressure drop and determination of the proper location for the catalyst in the air preheater to insure relatively uniform temperature distribution.

5.1.1 Overall System Design and Analysis

A schematic conceptual design of the pilot plant, showing design values of temperature, flowrate, and static pressure is presented in Figure 5.1. The flue gas sample point is located at an access hatch, upstream of the Unit 4 Air Preheater, in a section of the main duct through which there is flow at all times. This layout was approved by Valley Steam Plant personnel of the LADWP.

5.1.2 Catalyst Selection

Environics' previous small scale tests had indicated that a 0.3% platinum catalyst, supported on a high surface area wash-coat, would provide high activity for NOx removal in steam boiler flue gas, with a relatively low pressure drop through the catalyst bed. Thus, selection of the pilot plant catalyst was limited to such a catalyst, with the objective

of selecting the proper size support, to provide acceptable levels of activity and pressure drop for a utility plant. Testing was performed to determine pressure drop and activity characteristics of several catalyst/support configurations.

5.1.2.1 Pressure Drop Tests. Pressure drop experiments were performed on the various supports considered suitable for use in the pilot plant catalyst bed. The supports used were American Lava, 5 corrugations per inch, 10 mil wall thickness, split cell, and Hexcel A3-30, alumina, 30 holes per square inch, L3-50 lithium aluinum silicate, LAS, 50 holes per square inch, and L3-100, 100 holes per square inch. The experiments were performed on uncatalyzed support beds 6 inches square, at several depths (3", 6", 9" and 12" for 5/10-SC and L3-50) and at several flowrates ($\sim$ 180, 240, 350 SCFM for L3-100 and L3-50, and $\sim$ 240 and 350 SCFM for A3-30 and American Lava, 5 corrugations per inch. Figure 5.4 shows pressure drop (inches water column per inch bed depth) as a function of flow (SCFM) for the four supports. (A flow rate of 208 SCFM corresponds to the pilot plant design point.)

It appeared that the L3-50 support was desirable for the pilot plant application. A 12 inch deep catalyst bed, running at 208 SCFM per 6" square section, would impose a pressure drop (at 70°F) of 12 x 0.10" H_2O = 1.2" water column. Allowing a factor of 2 for the increase in pressure drop at 475°F, this gives a total hot pressure drop of 2.4" water column, slightly within that allowed on the schematic in Figure 5.1.

After pressure drop data on various possible catalyst supports were taken at 70°F, hot pressure drop tests at 475°F were made and evaluated.

Consideration of the hot pressure drop data at the flow rate indicated, also, that the best choice for catalyst support,

from the standpoint of pressure drop, would be L3-50. A 9-inch bed would impose a hot pressure drop of 2.1 inches H_2O, less than the 2.5 inches allowed in the overall design. If necessary, a 12 inch bed could be used, which would require only a slightly reduced hot gas flow rate at a pressure drop of 2.5 inches of water column.

5.1.2.2 Catalyst Activity Tests. To determine the suitability of Hexcel L3-50 as a support, samples of platinized L3-50 were obtained from Matthey Bishop for evaluation. These samples, assembled to form a catalyst bed 6" square x 12" deep, were placed in the 12000 SCFH test boiler reactor. The catalyst was tested with the following results.

A short test of one hour duration was conducted, at which time 95% NOx removal efficiency was observed. The system was shut down for the weekend, after which testing was resumed. During startup, the boiler was accidentally operated in a fuel-rich condition, resulting in incomplete combustion and deposition of carbon on the catalyst surface. Testing was continued for several days, on the theory that the carbon would burn or blow off spontaneously. However, at no time during those several days was removal efficiency in excess of 80% observed. The catalyst was therefore removed from the reactor and placed in a controlled atmosphere kiln, in which temperature and oxygen concentration were gradually increased to 750°F and 10% O_2, in order to "burn off" the carbon. The catalyst was then replaced in the reactor and tested continuously for four days. Immediately after this regeneration procedure, 94% NOx removal efficiency was observed. By the end of the four day test, this had declined to 92%.

In addition to these observations of removal efficiency, several samples of the catalyst were removed, and their BET surface area was measured. It was determined that the new

catalyst (prior to any testing) had a surface area of approximately 20 m^2/gm. The catalyst after carbon deposition (immediately prior to regeneration) had a surface area of approximately 11m^2/gm. The catalyst following regeneration had a surface measurement of 19 m^2/gm.

Ignoring the decrease in the removal efficiency which resulted from carbon deposition, this catalyzed L3-50 still appeared somewhat unstable, i.e., NOx removal efficiency decreased from 94 to 92% in only a few days. Furthermore, its BET surface area, when new, was considerably lower than that of catalyzed 8/8 SC (American Lava) (20 vs 37 m^2/gm). The conclusion was thus reached that, at that time, it could not be guaranteed that L3-50 could be catalyzed in a completely satisfactory manner. Therefore, it was decided to use catalyzed American Lava 8/8 SC as the pilot plant catalyst. (This catalyst is identical to that used in the long duration-small scale testing, and to the platinum catalyst used in the previous small scale work.)

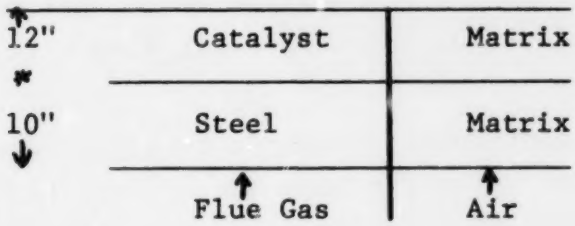
5.1.3 Catalyst Heat Transfer Analysis

A study of heat transfer in the air preheater wheel was performed, in order to estimate temperature distribution of air, gas and matrix in the wheel, and to determine whether catalyst temperature could be controlled as the power plant load changes. Details of the calculations are given below:

Basis of Calculations

1. Flue gas: Inlet temperature 700°F; flow rate: 50,000 SCFH.
2. Air: Inlet temperature; 70°F; flow rate: variable.
3. Wheel size: 7HKX-22: Air Preheater Company (standard speed: 3 RPM).
4. Co-current flow.

A schematic of the wheel, showing depth of hot end material (steel matrix) and cold end material (catalyst matrix) is presented below.



Calculations

The two most important variables which control the temperature distribution in the regenerative wheel are the speed of the wheel and the air flow rate. Therefore, calculations were carried out for three different cases:

Case I : RPM = 3; air flow = 50,000 SCFH/ft^2;
Case II : RPM = .41; air flow = 50,000 SCFH/ft^2;
Case III: RPM = 3; flue gas outlet temperature = 475°F. (Back calculate the air flow needed.)

The estimated matrix temperature, and average air and gas temperatures resulting from these calculations, are shown in Figure 5.2. In this graph, matrix temperature, as a function of axial position, is shown by lines for the three cases above. The indicated matrix temperatures are actually the temperature extremes experienced by the matrix, at the end of its passage through the flue gas (when it is hottest), and at the end of its passage through the air side (when it is coldest).

Conclusions

1. By changing the wheel speed from 3 RPM to .41 RPM, the matrix temperature excursions can be varied from less than

± 10°F to about ± 250°F. Therefore, this wheel can be used to simulate a wide range of real wheels, which are in use in utility steam power plants.

2. By changing air flow rate, catalyst matrix temperature can be controlled to any desired value, between approximately 700°F (air flow = 0) and 400°F (air flow = gas flow = 50,000 SCFH/ft^2).

3. In particular, catalyst temperature can be controlled to the desired value for peak NOx removal efficiency, i.e., 475°F, by using an air flow rate of 36,000 SCFH/ft^2. At the standard 7 HKX-22 wheel speed of 3 RPM, temperature excursion of the catalyst matrix under these conditions will be <20°F.

5.2 CONTROL SYSTEM DESIGN

The pilot plant control system was designed to accomplish four functions: (1) inlet NH_3 concentration control, (2) flue gas temperature control, (3) flue gas flow measurement and control, and (4) air preheater inlet pressure control. The various functions are described below.

5.2.1 Inlet Ammonia Concentration

Three methods to control NH_3 concentration were considered. The three methods are:

Method 1: Control NH_3 flow as a function of inlet flow rate and NOx concentration.

Method 2: Control NH_3 flow to a constant concentration in the inlet stream.

Method 3: Control NH_3 flow to minimize the total NOx + NH_3 concentration in the exit stream.

Process control equipment manufacturers were contacted to determine the nature of the equipment required for the control

system. Method 3 would be the most desirable. However, the hardware for this system is relatively sophisticated and was not readily available.

It was thus decided to simulate the control method 3 to control the ammonia injection rate manually, using a hand valve and calibrated rotameter.

5.2.2 Flue Gas Control System

Flue gas temperature is controlled with a temperature controller measuring gas temperature on the air preheater flue gas exit stream. The controller adjusts the position of the air flow control damper operator to provide the air flow rate required to maintain the set temperature.

In addition, high temperature protection is provided by a control loop separate from the above temperature controller. A separate high limit controller monitors the temperature of the exit gas. If this temperature rises above a set limit (e.g., 650°F), the controller will turn off the flue gas fan, close the flue gas inlet damper, and close a solenoid in the NH_3 supply line. This controller requires manual reset before operation of the system can be resumed.

5.2.3 Flue Gas Flow Control System

The flue gas inlet damper is controlled by a modutrol motor and an open/close switch. The damper is held full open or full closed in response to this switch position, and a pressure gauge on the control panel indicates whether the damper is open or closed.

Flue gas flow rate is measured with a venturi flowmeter. The pressure drop across the venturi is measured and read on a Dwyer Photohelic differential pressure gauge on the control panel.

Flue gas flow rate is controlled by setting the flue gas ID fan damper position using a modutrol motor. This control can be performed either manually, with a panel switch, or automatically with switches in the differential pressure gauge. A panel meter indicates the position of the damper (0 to 100% open).

5.2.4 Air Preheater Inlet Pressure Drop Control

In order to prevent leakage from the preheater air side to the preheater gas side, or vice versa, it is necessary to equalize the pressure at the air inlet with that at the gas inlet. This is accomplished by measuring the pressure drop between the air and gas inlets with a differential pressure switch. The output of this switch is used to open or close the inlet air pressure damper, through a modutrol motor, to achieve the desired condition.

5.2.5 Installation and Safety Consideration

A safety system will shutdown the pilot plant if the set point of the high limit temperature controller is exceeded or if the electrical signal that the plant is firing fuel oil is received. In either case, the shutoff damper will close, the ammonia flow solenoid valve will close, the flue gas ID fan will stop, and a signal will be sent to an annunciator in Unit 4 control room. Valley Steam Plant personnel were instructed to call Environics personnel if such a shutdown occurred during a period of unattended operation.

Pressure taps are ¼" swagelok fillings brazed to the outside of the duct and flush with the duct wall on the inside. Sample taps are ¼" stainless tubes extending to the center of the duct. Thermocouples are 1/8" diameter Type J, extending to the center of the duct. Twelve Type J thermocouples are

provided, 9 cemented to the catalyst surface and 3 extending into the gas stream.

5.3 AMMONIA INJECTION SYSTEM DESIGN

5.3.1 Laboratory Tests

Tests on the NH_3 injection system were conducted with the small scale test boiler.

The test apparatus size was 6" square; a flow of 50,000 SCFH/ft^2 was maintained to simulate the flow regime of the pilot plant. Samples were taken from the center of each square in the 9 point sample grid, and analyzed for NH_3 by the oxidation method described below.

NH_3 was injected through a conical spray pattern orifice, manufactured by Spraying Systems Co., catalog #1/8 BX0.5, 316SS. NH_3 concentration was determined at each point of the nine point sampling array, 14 duct diameters downstream of the nozzle.

Results of three average complete sets of data are shown in Figure 5.3. Mixing was fairly uniform, except for a low average concentration in the bottom row.

On the theory that the injection nozzle was tilted slightly upward, a fourth scan was made after pointing the nozzle 20 degrees downward of its initial position. These results are shown in Figure 5.4. For this test, the concentration in the bottom row was higher than that in the top two.

The following conclusions were reached from these tests:

1. Injection through a simple nozzle will provide satisfactory mixing for use in the pilot plant.
2. Nozzle alignment is somewhat critical, in order to obtain uniform mixing.

3. Use of mixing devices or inert gas dilution will not be necessary to provide adequate mixing.

5.3.2 Pilot Plant Injection System

On the basis of the small scale tests, the NH_3 injection system for the utility pilot plant was designed with sixteen injection nozzles, one for each 6" square in the 24" square inlet duct. This nozzle array was located as far upstream of the catalyst as possible, to take maximum advantage of the available mixing length.

5.4 ANALYTICAL TECHNIQUES

The following analytic techniques were selected for use in the pilot plant program.

5.4.1 NOx Analysis

Oxides of nitrogen were analyzed with a Beckman Model 915 chemiluminescence analyzer.

5.4.2 Ammonia Analysis

Ammonia analysis was performed by two techniques. The bulk of the analyses was performed using catalytic oxidation of ammonia to NO, followed by analysis of the product NO in the chemiluminescence analyzer. Periodic checks were made by wet chemistry. These techniques are described below.

Catalytic Oxidation of NH_3

The reactor used to analyze for NH_3 by oxidation to NO is described below.

The oxidizer uses 50 cm^3 of catalyst (2% Pt on gamma Al_2O_3 pellets, 1/8" diameter) which, with a gas flow of 10

SCFH, gives a space velocity of 6,000 hr^{-1}. The oxidizer converts NH_3 to NO by the reaction:

$$4NH_3 + 5O_x \text{ -----> } 4NO + 6H_2O.$$

In order to determine the temperature required for 100% conversion of NH_3 to NO, the following tests were performed. Premixed NH_3/N_2 gas (400 ppm) was passed through a calibrated rotameter and then mixed with dry air to achieve a 2% O_2 concentration, before entering the reactor. The ammonia concentration (NH_3) was determined by the standard wet chemical method. The product (NO) of this reaction was analyzed by a Beckman NDIR analyzer.

By varying the temperature of the catalyst bed, a curve (Figure 5.5) showing conversion of NH_3 to NO, as a function of temperature, was obtained. The results indicated that 100% conversion was obtained at a temperature of 1200°F. The same result was also obtained by Klimisch and Barnes (Environmental Science and Technology, June 6, 1972).

It was concluded that by operating the oxidizer at the proper temperature (1200°F), 100% conversion of NH_3 to NO could be obtained, allowing rapid response continuous analysis for NH_3, using a conventional NO analyzer.

This reactor was used in the NH_3 mixing studies, allowing the opportunity to check conversion efficiency over a longer period of time than was done in the temperature studies, with satisfactory results.

5.4.3 Oxygen Analysis

Oxygen is analyzed in the pilot plant using a Beckman Model 715 Process O_2 Analyzer.

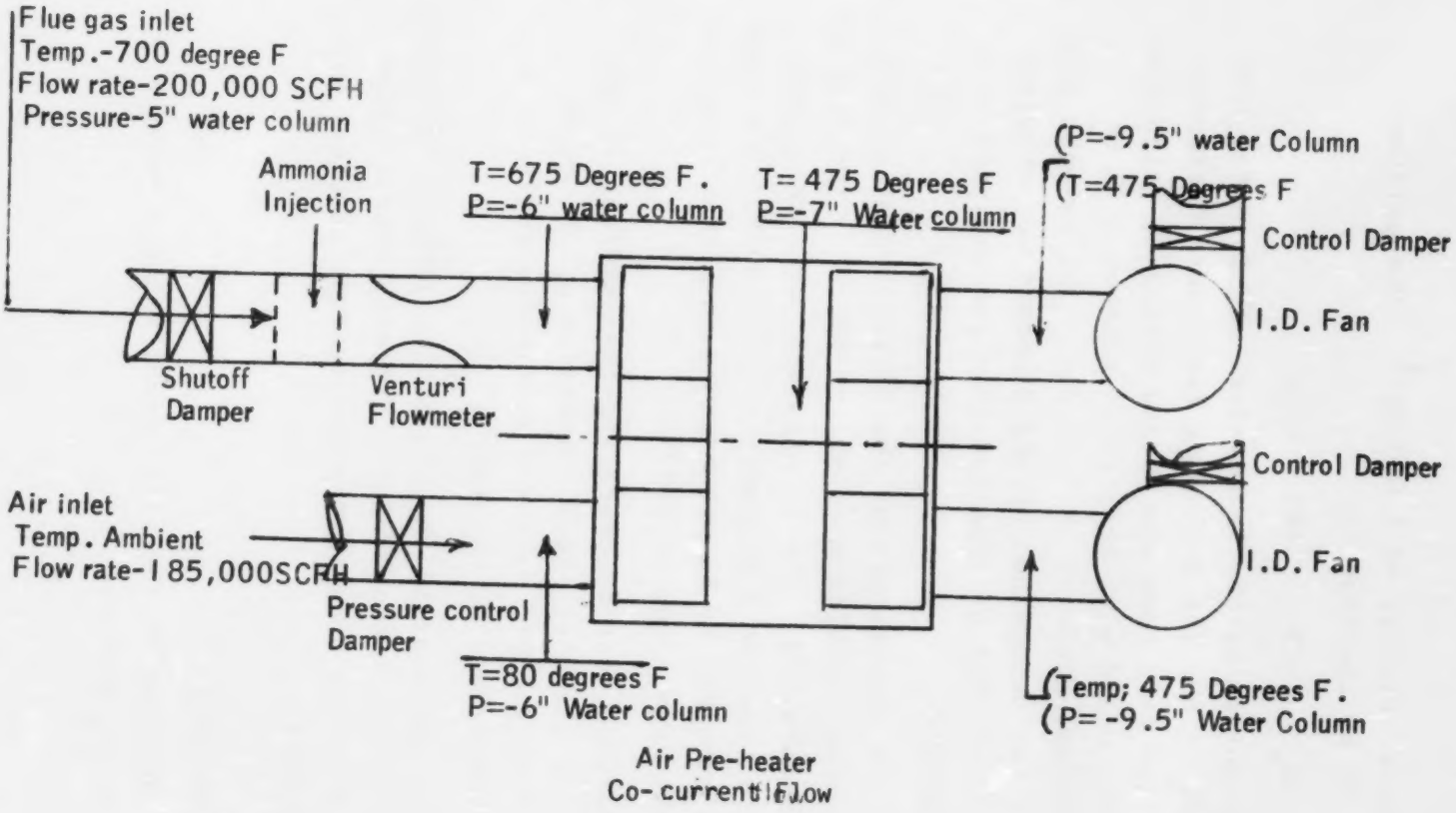


Figure 5.1 Utility Pilot Plant Schematic

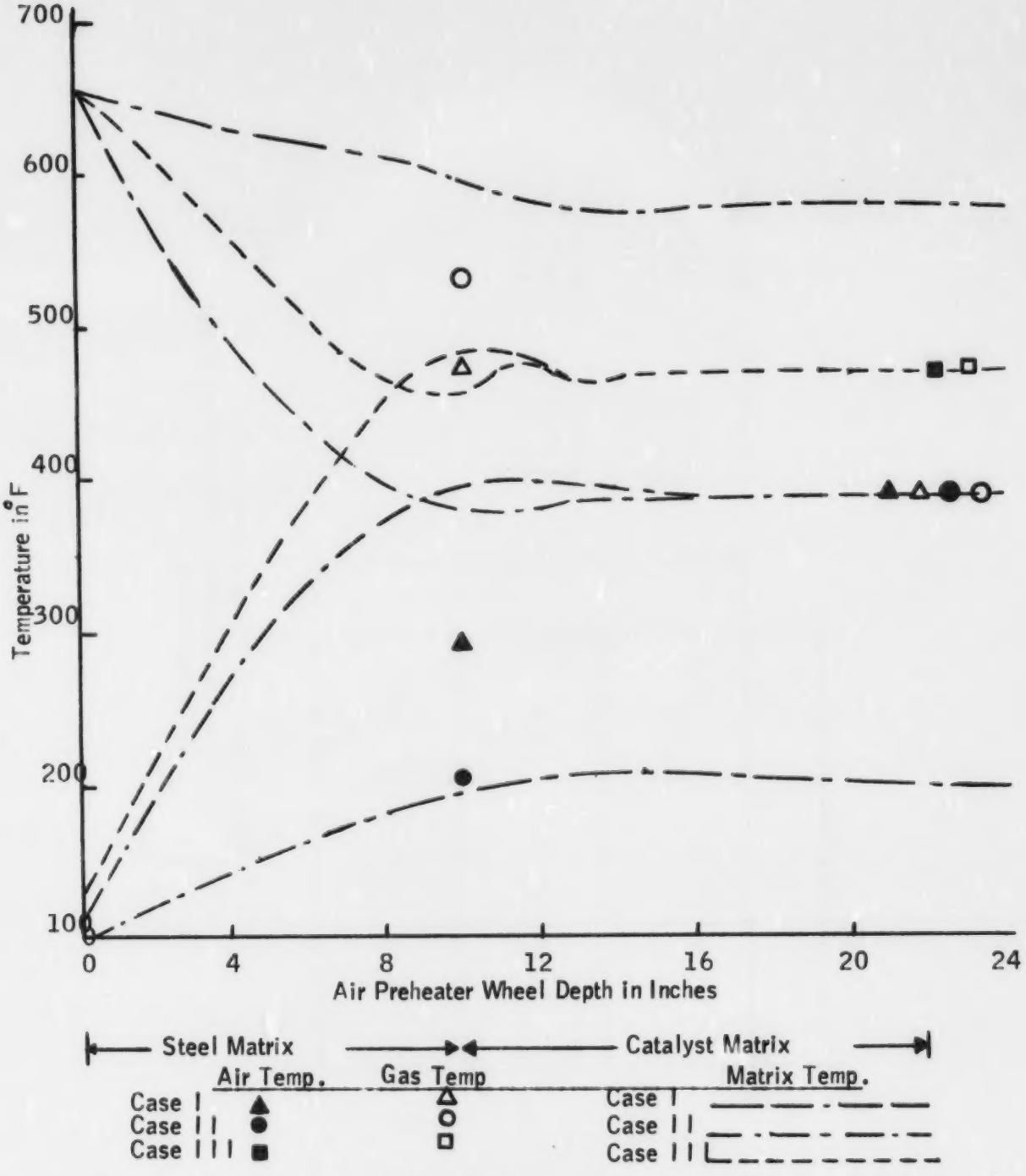
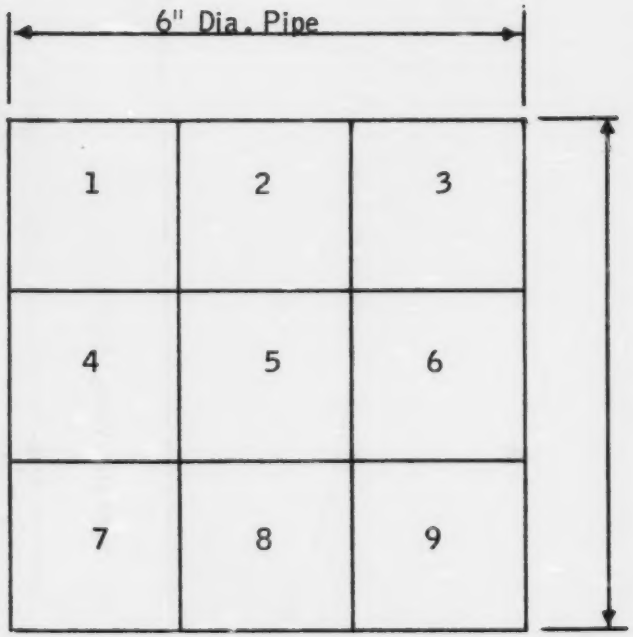


Figure 5.2 Calculated Temp. Profile
Air Preheater 7HKX-22



Average (NH_3 + NO) in ppm	
1+2+3	428
4+5+6	422
7+8+9	396
1+4+7	415
2+5+8	392
3+6+9	420

Figure 5.3 Ammonia Mixing Test Results
Injection Nozzle Zero Degree to Horizontal
Boiler NOx=45 ppm

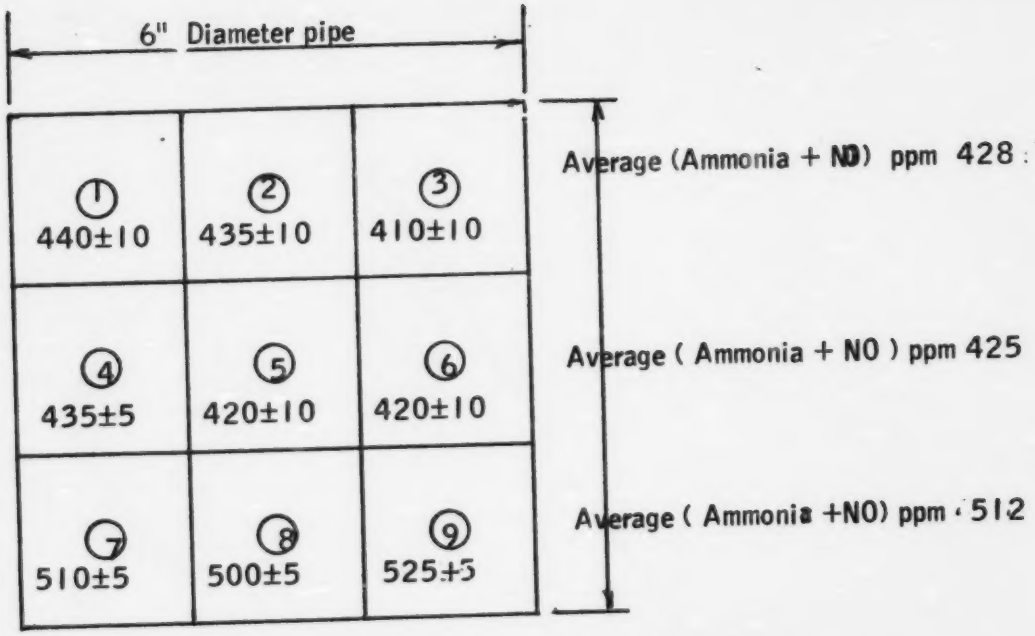


Figure 5.4 Ammonia Mixing Test Results
Ammonia Injection Nozzle negative Angle 20
Boiler Ambient NOx=45ppm

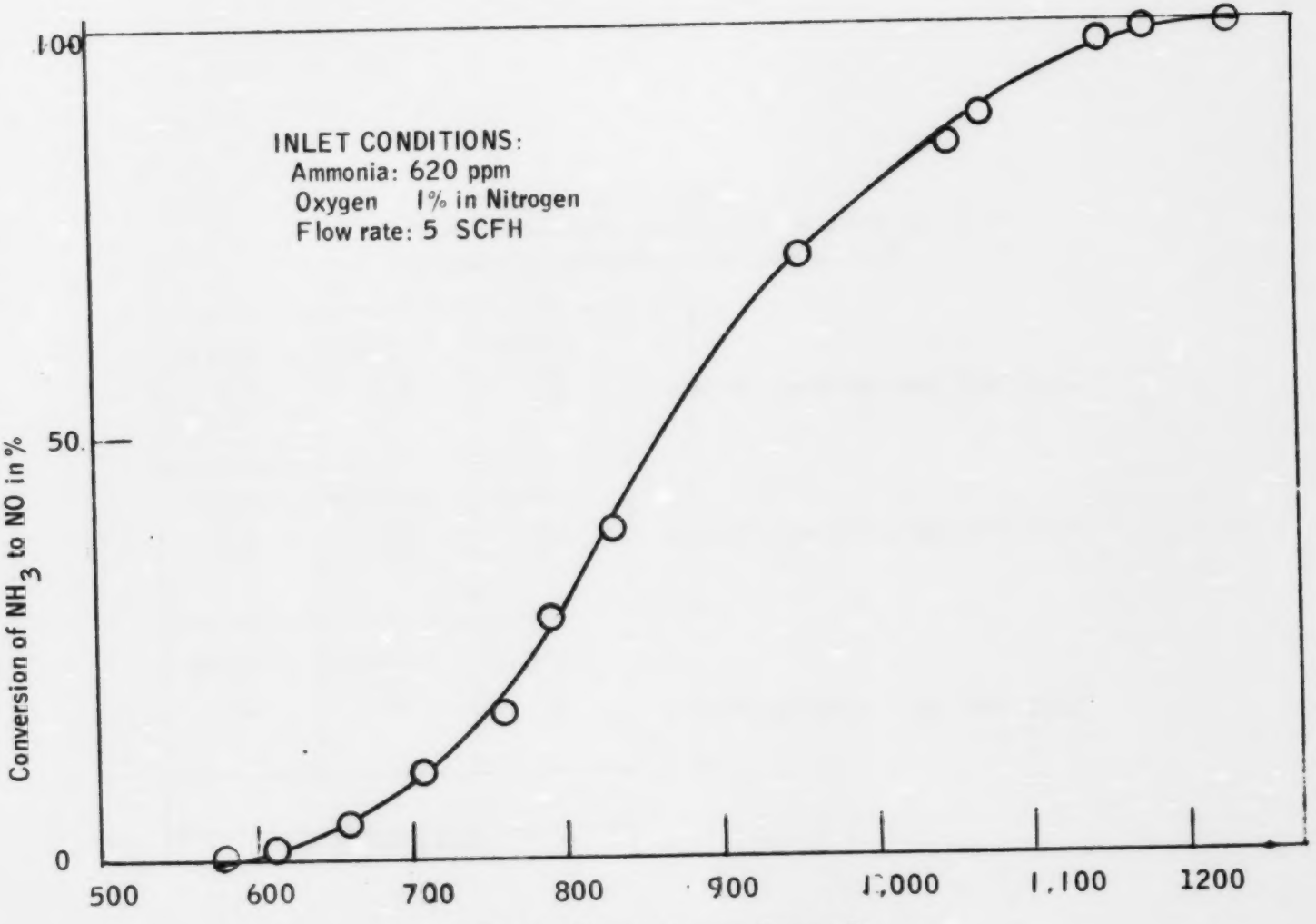


Figure 5.5

Conversion Efficiency vs. Temperature for NH_3 Oxidizer

SECTION 6

LABORATORY PILOT PLANT TESTING

During the design stage of the utility pilot plant program, laboratory pilot plant testing was performed to validate results of previous work and to provide previously unavailable information on catalyst life and possible formation of trace exhaust products.

These tests were performed using a 550,000 BTU/hour gas fired boiler, equipped with an ID fan to direct a portion of the flue gas through the catalyst bed. Flow through the reactor was measured using an orifice plate. Analysis for NOx was done with either a Beckman Model 315 BL Non Dispersive Infrared Analyzer (sensitive to NO only) or a Beckman Model 971 Chemiluminescence Analyzer. Oxygen concentration was determined with a Beckman Model 715 Process O_2 Monitor. Ammonia concentration was determined using the catalytic and wet chemical techniques previously described.

6.1 LONG DURATION TESTING

Long duration testing was conducted to provide an estimate of catalyst performance under exposure to actual gas fired flue gas over a long period of time. These tests were also expected to provide warning of any catalyst deterioration that might occur in the utility pilot plant, so that steps could be taken to investigate appropriate regeneration techniques.

Long duration testing was conducted for a total time of 175 days. Conditions during the test were as follows:

Temperature:	490 to 500°F
NH_3 concentration (inlet):	200 to 250 ppm
NOx concentration (inlet):	240 ppm

Flue gas O_2 concentration:	5%
Space Velocity:	50,000 v/v/hr
Catalyst:	0.3% Pt on 8/8 SC support (MBI), 4"x4"x3" thick.

Results for the entire test period using weekly averages of the data are shown in Figure 6.1. Figure 6.1 shows NOx removal efficiency, total removal efficiency and temperature. It can be seen that NOx removal efficiency remained approximately 95% and total removal efficiency remained between 80-85% for the entire test.

During the test period on days 151-154, it was discovered that the (NOx + NH_3) exit concentration was unusually high (100 ppm). It is believed that this was due to an upset in boiler operation (accidental shutdown of the induced draft fan) which resulted in poor combustion and exposure of the catalyst to carbon particles. After the condition had persisted for several days, and could not be corrected by changing NH_3 concentration, it was decided to remove and attempt to regenerate the catalyst. (It was noted during disassembly of the converter that the orifice plate, immediately downstream of the catalyst, was indeed covered with fine soot.)

The regeneration procedure was similar to that used previously, with the L3-50 catalyst. The catalyst was placed in a kiln, through which air was passed at a low flow rate (5 SCFH), and held at 400°F for 30 minutes. The temperature was gradually increased to 700°F, over a period of 90 minutes, and maintained there for 16 hours.

After this treatment, the catalyst was replaced in the converter, and testing resumed. Activity was restored to previous levels; the observed (NOx + NH_3) exit concentration after regeneration was again approximately 40 ppm.

6.2 PARAMETRIC TESTING

Parametric testing was performed, using the same gas fired boiler and catalyst as in the long duration testing, to determine the effects of space velocity, inlet NH_3 concentration, and temperature upon removal efficiency. This testing was desired for two reasons: (1) to extend previous small scale data and (2) to provide scale-up information upon which design and evaluation of the utility pilot plant could be based.

6.2.1 Effect of Inlet NH_3 Concentration

Two parametric studies of the effect of inlet NH_3 concentration on removal efficiency and exit NH_3 concentration were performed. Figure 6.2 shows the results at 490°F. In general, if inlet ammonia injection exceeded 250 ppm, the NOx removal efficiency did not increase and the exit ammonia concentration increased.

6.2.2 Effect of Space Velocity

Experiments were performed to determine the effect of space velocity on NOx removal efficiency. Results, shown in Figure 6.3, are similar to what has been observed previously in smaller scale testing. Removal efficiency decreases with increasing space velocity; however, between 50,000 and 75,000 v/v/hr this effect is not very pronounced.

6.2.3 Effect of Temperature

Figure 6.4 shows results of experiments performed to determine the effect of temperature on NOx removal efficiency. Data is presented at 470 and 500°F, showing NOx removal efficiency and NH_3 exit concentration as a function of NH_3/NOx ratio in the inlet gas. Removal efficiency is relatively

unchanged by temperature in this region; however, NH_3 exit concentration decreases quite rapidly with increasing temperature. Unfortunately, it was not possible to run any experiments at higher temperatures. Numerous attempts to both increase fuel flow and to plug boiler tubes produced little or no temperature increase. However, previous data indicate that at 525°F, removal efficiency would be only slightly lower than at 500°F and NH_3 exit concentration would be significantly lower.

6.3 TRACE PRODUCTS ANALYSIS

It was desired to determine whether any undesirable trace products were formed as byproducts of the NOx/NH_3 reaction over the platinum catalyst. To accomplish this, analysis of three samples of the exit gas from the long duration test catalyst was performed by TRW, Inc. Results are shown in Table 6.1.

Several comments may be made about this data:

(1) Environics' in-house analysis of the exit gas, conducted immediately before these three samples were taken, found 14 ppm NO, zero ppm NO_2 and 13 ppm NH_3. The difference between our analysis and TRW's for NO and NO_2 cannot be explained readily. The difference in the NH_3 analysis is probably due to adsorption of the NH_3 in the water condensed in the sample flasks. This water was not analyzed.

(2) The fact that no N_2O or HCN was found indicates that these compounds are not produced by this reaction. Although this data is not conclusive, it was felt to be sufficient, and the trace products investigation was considered complete.

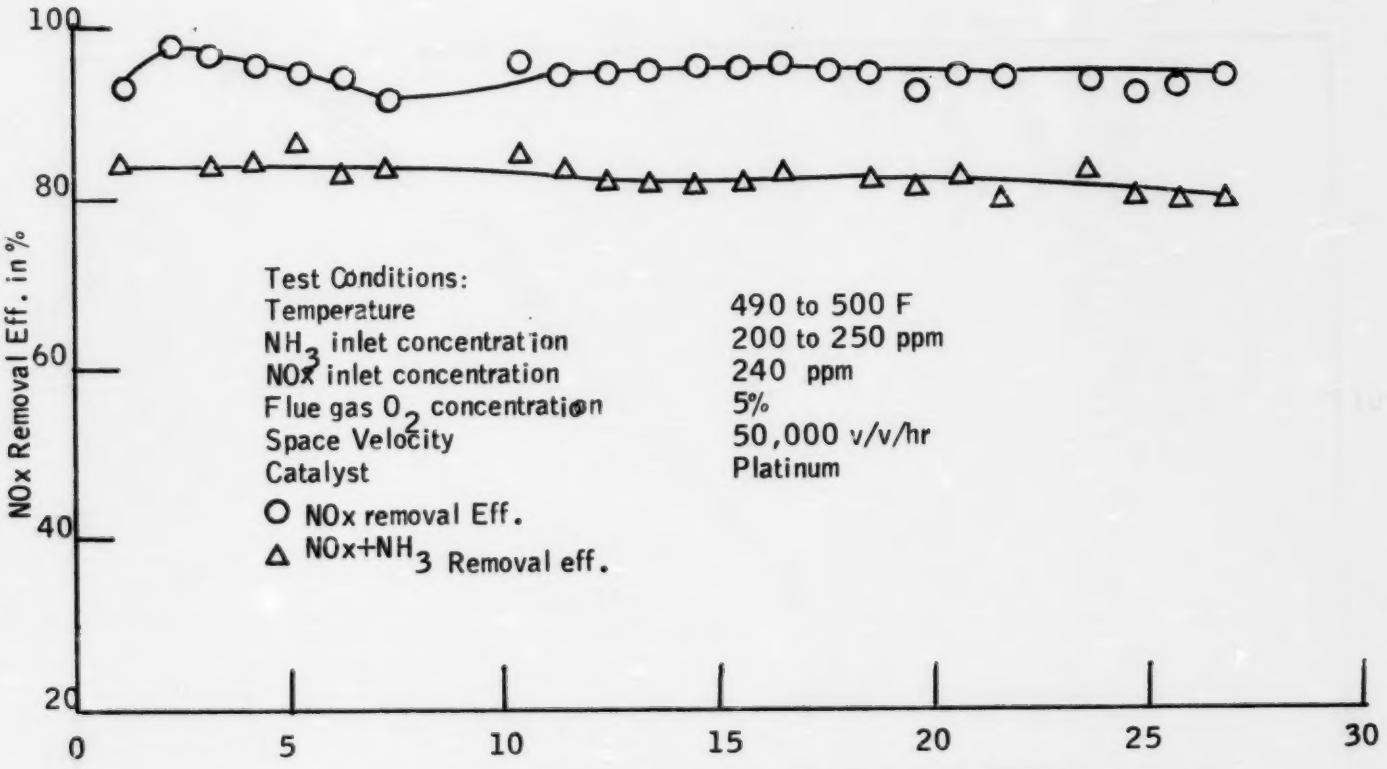


Figure 6.1 Catalyst Life Test in Weeks

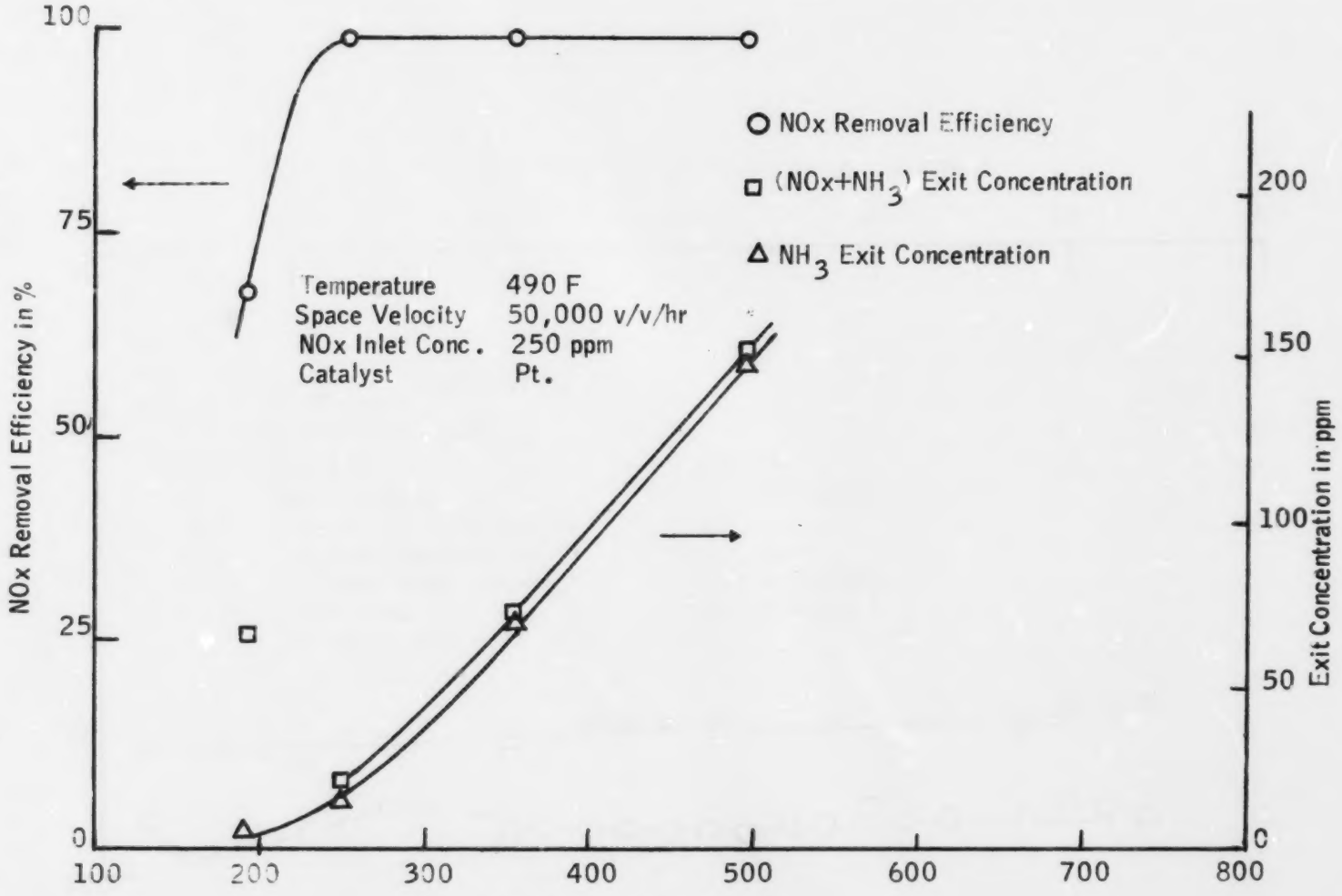


Figure 6.2 Effect of NH_3 on NOx Removal Efficiency

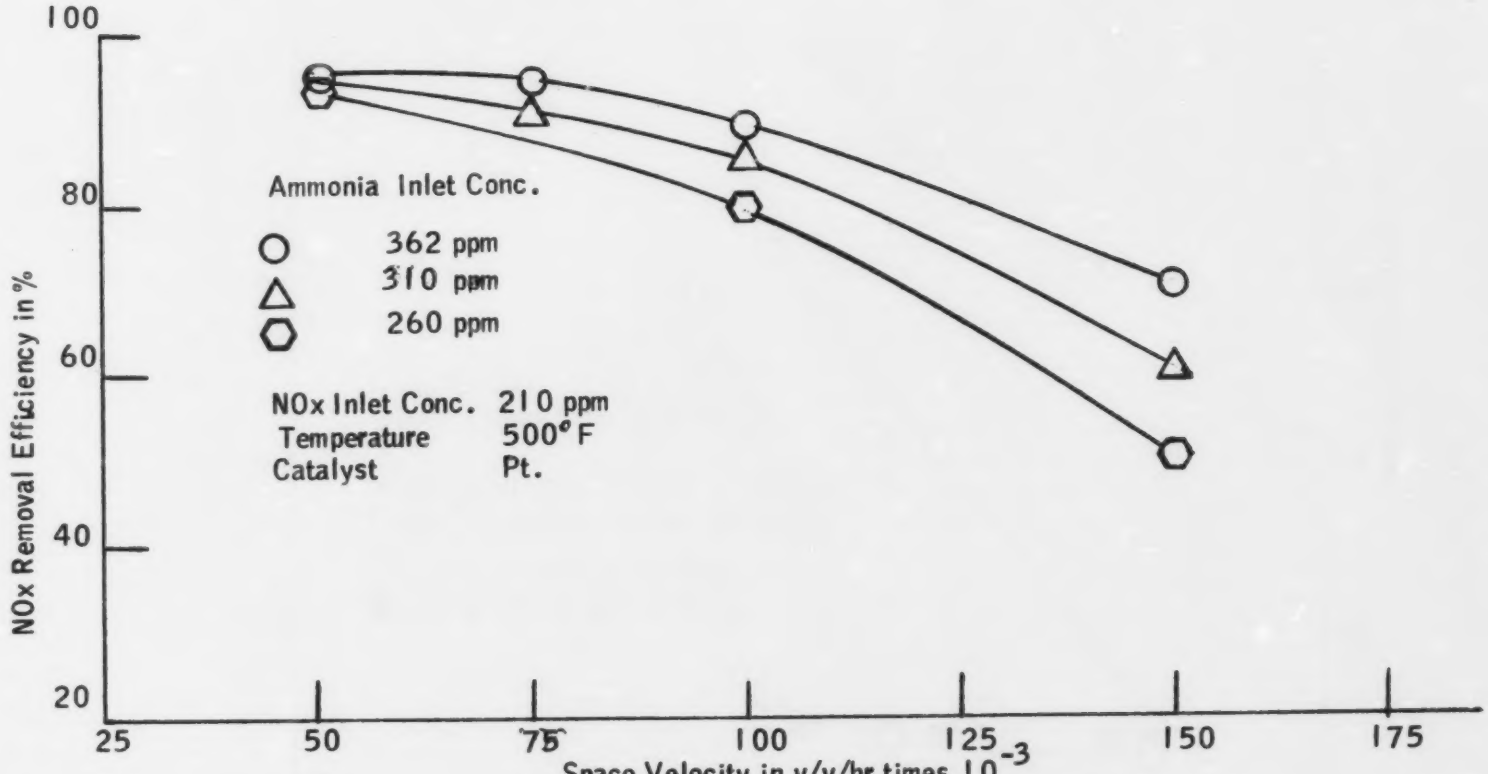


Figure 6.3 Effect of Space Velocity on Removal Eff.

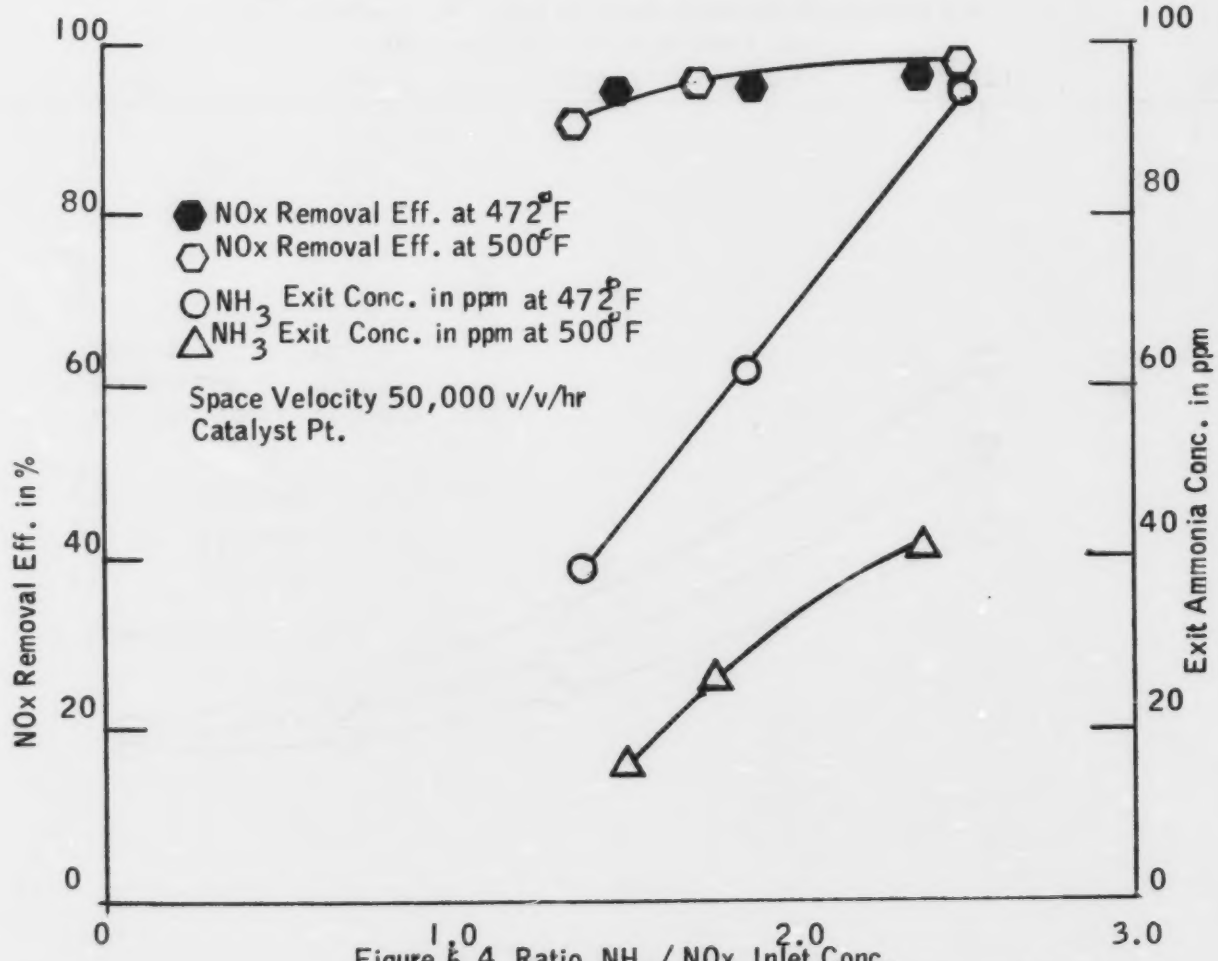


Figure 6.4 Ratio NH_3/ NOx Inlet Conc.
Effect of Temperature and NH_3/NO Ratio on Removal Eff.

Table 6.1

Trace Products Analysis by TRW

Sample #	NO ppm	NO_2 ppm	NH_3 ppm	N_2O ppm	HCN ppm	CO_2 %	O_2 %
1	32	15	0	0	0	6	5
2	32	0	0	0	0	6	5
3	36	6	0	0	0	6	5

SECTION 7

PILOT PLANT INSTALLATION AND CHECKOUT

7.1 INSTALLATION

Installation of the utility pilot plant was performed during the months of March and April, 1973. Los Angeles Department of Water and Power personnel were employed to lift the air preheater to the work level and to position it in its proper location. Environics' personnel performed the remainder of the installation work; i.e., installation of the air and gas ID fans, ductwork, flow control dampers, ammonia injection system, control panel, and control system components. A photograph of the completed pilot plant is shown in Figures 7.1 and 7.2.

7.2 INITIAL CHECKOUT

Initial system checkout work was performed before the catalyst baskets were installed in the air preheater wheel. Details of these tests are described below.

Sampling System. The sample handling system was found to function properly, except that it was felt desirable to reverse the positions of the Hankison dryer and sample pumps, so that the pumps are upstream of the dryer. After installation of normally closed solenoid valves on the drain ports of the dryer, the system automatically blows out condensed water, when the sampling system is in the N_2 position.

It was also found necessary to modify the sampling system by adding heating tape to the inlet and exit samples manifolds, in order to eliminate water condensation and thus provide stable NH_3 analyses through the oxidizers.

Pressure Readout System. Pressure readout is provided at the system inlet, gas flowmeter and preheater air-gas inlet. The Dwyer magnehelic gauges were found to function satisfactorily; however, it was felt desirable to replace the 0-1" W.C. gauges provided for the flowmeter and air-gas inlet with 0-2" W.C. and 5-0-5" W.C., respectively, to provide a wider range for readout during startup and off-normal conditions.

Temperature Readout System. The 24 point temperature recorder was found to function satisfactorily.

Manual Damper Control Systems. Results of tests of the various manual control loops are presented below.

Shutoff Damper: A stiffer linkage rod was required to obtain satisfactory operation.

Air Pressure Damper: A stiffer linkage rod was required to obtain satisfactory operation.

Gas Control Damper: Satisfactory operation was obtained with no significant changes.

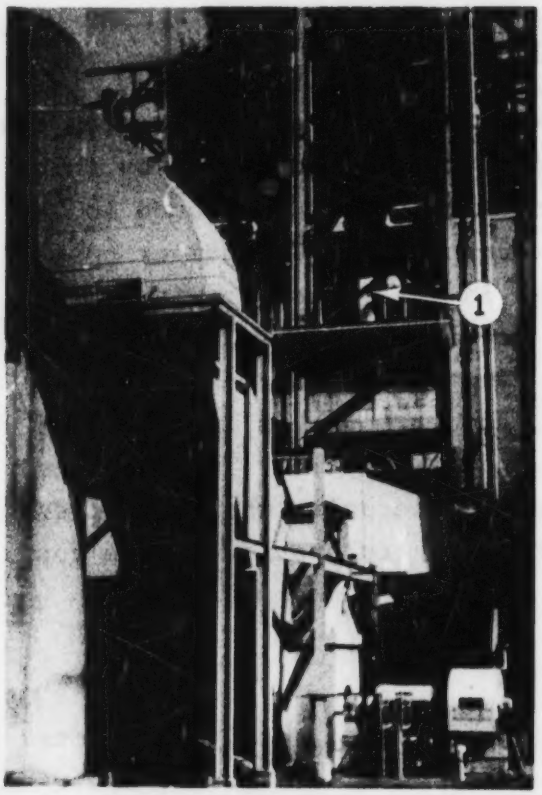
Temperature Control System. Initially, severe overshoot and oscillation was observed in the temperature control system. Adjustment of the temperature controller proportional band and auto reset potentiometers improved, but did not completely eliminate the problem. Analysis of the situation indicated that there were two other factors contributing to this problem. The first was the high degree of leakage within the air preheater. This was due to the absence of the cold end basket covers; when these covers were installed, the leakage was significantly decreased and temperature control was improved. The second trouble source was the lack of automatic control of the air inlet pressure damper. An automatic control system for this damper was installed and tested.

Air Pressure Damper Automatic Control System. To achieve automatic control of the air pressure damper, a Dwyer 1640

differential pressure switch was installed and the system was rewired to allow the damper to be controlled either by this switch or manually. This automatic control was found to be unstable, due to the delay between a change in damper position and the resulting response in the pressure switch diaphragm. It was decided to decrease the speed of the damper motor, by installing a cam timer in series with the pressure switch, such that the damper motor is actuated for only one second in each 10 second time cycle. This had the effect of slowing damper travel to the point where the pressure switch could respond fast enough to provide stable control as boiler load or temperature setpoint changed.

7.3 FINAL CHECKOUT

To accomplish the final checkout, the catalyst baskets were loaded into the air preheater. Asbestos gasket material was used to prevent gas bypassing the catalyst. Initial testing with catalyst in place indicated that the design flow rate of 180,000 SCFH could not be reached. Maximum possible flow was 115,000 SCFH; a flow of 157,000 SCFH is required for 50,000 v/v/hr space velocity. Flowrate was subsequently increased to 140,000 SCFH (45000 hr^{-1} space velocity) by increasing the speed of the flue gas ID fan.



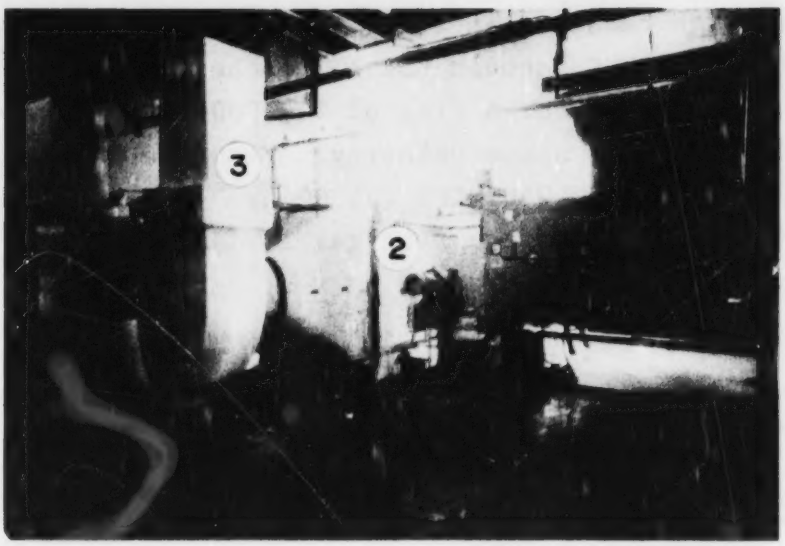
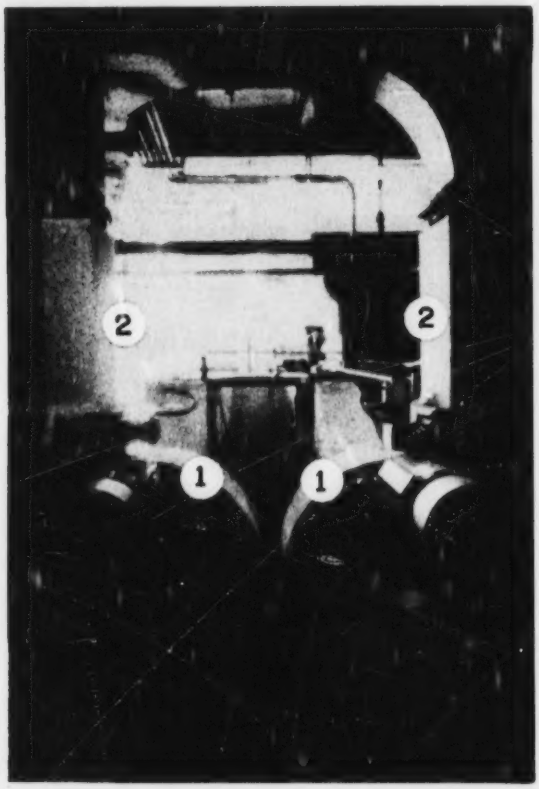


Figure 7.1 Utility Pilot Plant
1-Inlet Duct From Boiler Economizer
2-Catalytic Converter & Air pre-heater
3-Exhaust Ducts



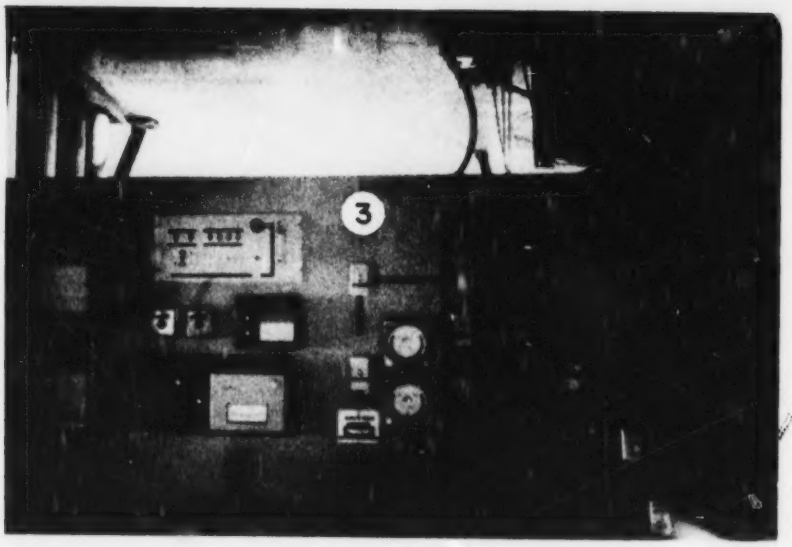


Figure 7.2 Utility Pilot Plant
1-Exhaust I.D. Fans 2-Exhaust Duct 3-Instrumentation Panel

SECTION 8

GAS FIRED PILOT PLANT TESTING

UTILITY PILOT PLANT

The bulk of the utility pilot plant testing was performed during periods when the steam plant was burning gas fuel. Following this work, tests on oil fuel were conducted. The results of the gas-fired tests are described below; results of the oil fired tests are described in Section 9.

Gas fired testing was conducted in two areas: parametric tests, with the objective of determining optimum operating conditions for the utility pilot plant, and long duration tests, with the objective of determining the long term performance of the catalyst under extended exposure to gas fired flue gas. In addition, tests were performed to determine whether the NH_3 was sufficiently well mixed with the bulk of the flue gas prior to entering the air preheater, which contained the catalyst bed.

8.1 UTILITY PILOT PLANT PARAMETRIC TESTS

Tests were performed to determine the effects of temperature and inlet NH_3 concentration on NOx removal efficiency and total removal efficiency. NOx removal efficiency and (NOx + NH_3) exit concentration, as a function of temperature, at a constant inlet NH_3 concentration were measured.

Inlet NH_3 concentrations of 100, 200, 300, 400 and 500 ppm were probed. Space velocity for these tests was 45,000 hr^{-1}.

This data indicate that 200 ppm ammonia is sufficient to obtain 85% removal efficiency at the typical NOx inlet concentrations encountered (125 ppm) and that higher ammonia

concentrations do not significantly improve this removal efficiency, but only increase the amount of unreacted ammonia in the exit flue gas. The data also indicates that the greatest NOx reduction is obtained at exit flue gas temperatures of 490 - 500°F. Figure 8.1 gives the optimum results.

8.2 PILOT PLANT NH_3 MIXING TESTS

Tests were also made to determine NH_3 distribution at the inlet of the air preheater. Each of the six probes in the inlet sample manifold was sampled individually to determine NH_3 concentration. Results are shown in Figure 8.2. Average NH_3 concentration at that time was 277 ppm. The difference between the various sample points was less than $\pm$ 6%.

8.3 PILOT PLANT LONG DURATION TESTS

During operation of the pilot plants, a total time of 2021 hours on gas firing was achieved. This testing was conducted during the period from July 1973 through July 1974. Although the pilot plant was operated whenever conditions of operation of Valley Steam Plant Unit 4 permitted, there were numerous shutdown periods. The pilot plant was shut down for any of the following reasons: oil firing on Unit 4, unscheduled repairs to Unit 4 (tube leaks, etc.), scheduled maintenance to Unit 4, low load on Unit 4, and a strike by Valley Steam Plant personnel.

Results of the gas fired duration tests are presented in Figures 8.3 through 8.5. Each figure shows NOx removal efficiency, gas temperature (at the catalyst exit) and NH_3 inlet concentration, as a function of catalyst operating time in hours. These figures include data from the parametric tests as well as the long duration tests at optimum conditions. It can be seen that, over the duration of this testing, no

indication of catalyst deterioration was observed, i.e., at optimum conditions, NOx removal efficiency of 75-90% was achieved throughout the test period.

It has been shown by tests performed by Environics prior to the contract initiation and confirmed by the long duration laboratory testing during the contract performance that NOx removal efficiency is sensitive to the temperature of the flue gas and the amount of ammonia injected. Figure 8.1 shows that the maximum removal efficiency occurs at a temperature between 490° and 500°F. at an ammonia injection of 200 ppm. Increase of the ammonia injection will increase the NOx removal efficiency; however, the effluent ammonia may also be increased. Therefore, the utility pilot plant long duration testing tried to probe the effects of temperature and ammonia injection while other operating conditions were maintained constant for a relatively long time.

Figures 8.3 through 8.5 show that:

(1) For approximately 700 hours the NOx removal efficiency was essentially 85% at temperatures between 460 and 470°F. Ammonia injection was maintained at 200 ppm. Space velocity obtained was 45,000 v/v/hr.

(2) Attempts were made to decrease ammonia injection to 125 ppm at lower flue gas temperatures. Under these conditions, the NOx removal efficiency decreased to approximately 60%.

(3) Attempts were also made to increase ammonia injection at higher flue gas temperatures. Under the conditions of 500 ppm ammonia injection and flue gas temperature at 530°F, the NOx removal efficiency increased to slightly higher than 85%. Although there was no measurement of effluent ammonia, it is believed that the effluent ammonia did increase.

(4) When the system operating parameters were returned to the initial values, namely, 200 ppm ammonia injection and flue gas temperature between 460-490°F, the NOx removal efficiency did return to approximately 85%, previously obtained.

It appears that the results of the utility pilot plant long duration testing did meet the objective of the program. It has demonstrated that: (1) the ammonia reduction system can remove NOx at 85% efficiency; (2) the catalyst did have a life more than 2000 hrs without deterioration; and (3) it can operate at a space velocity of 45,000 v/v/hr for more than 2000 hours.

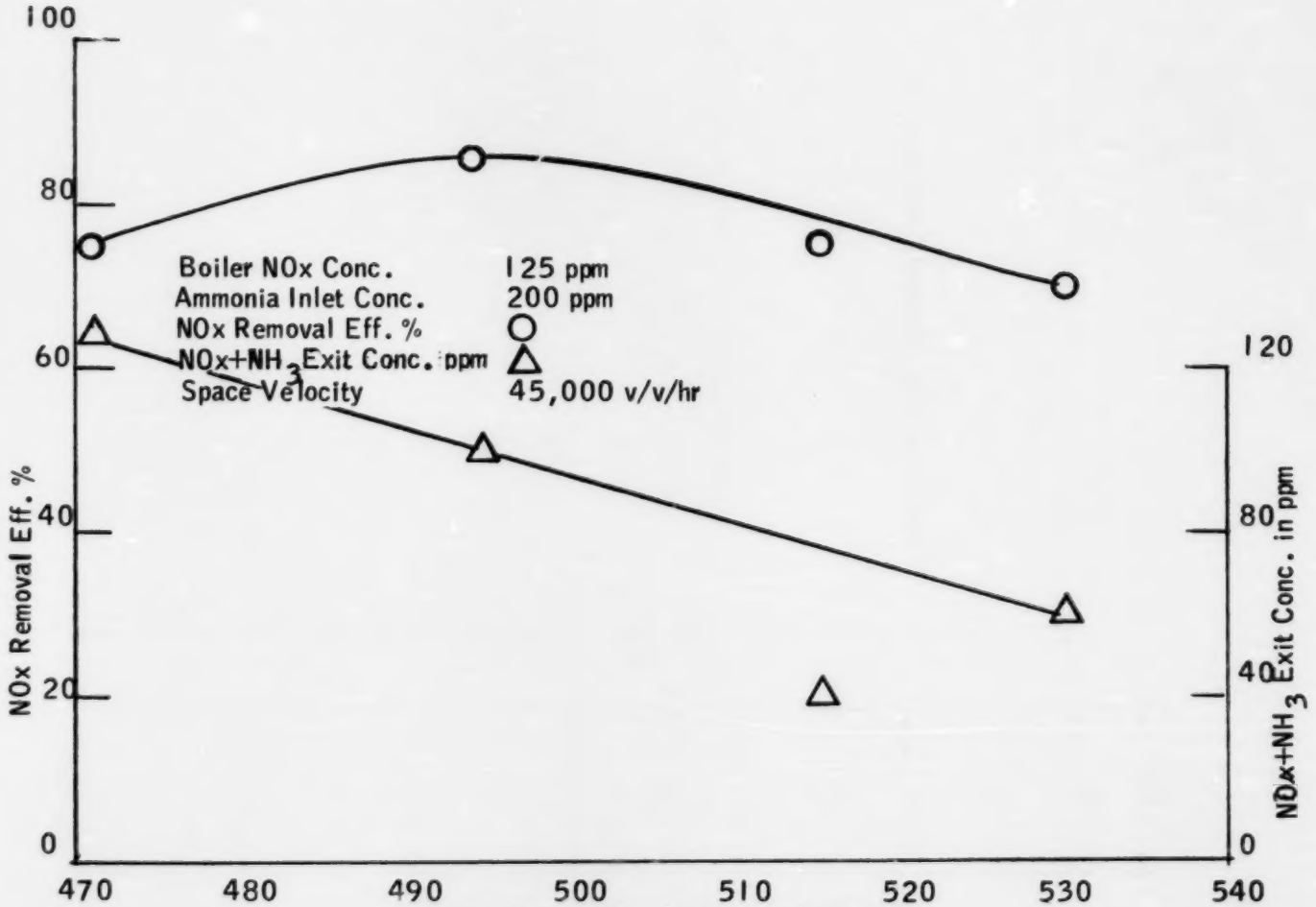
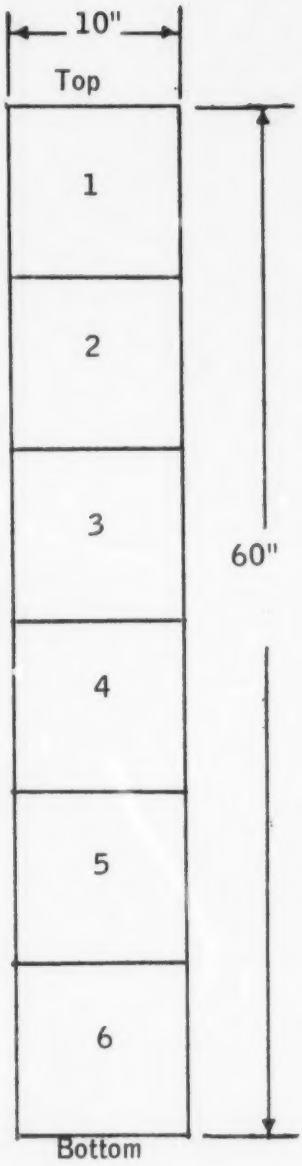


Figure 8.1 Optimum Test Conditions-Utility Pilot Plant



Duct Dimension and Probe Arrangement

Probe Location	NH_3 Conc. in ppm
1	263
2	273
3	286
4	290
5	291
6	258

Figure 8.2
Inlet NH_3 Concentration Distribution

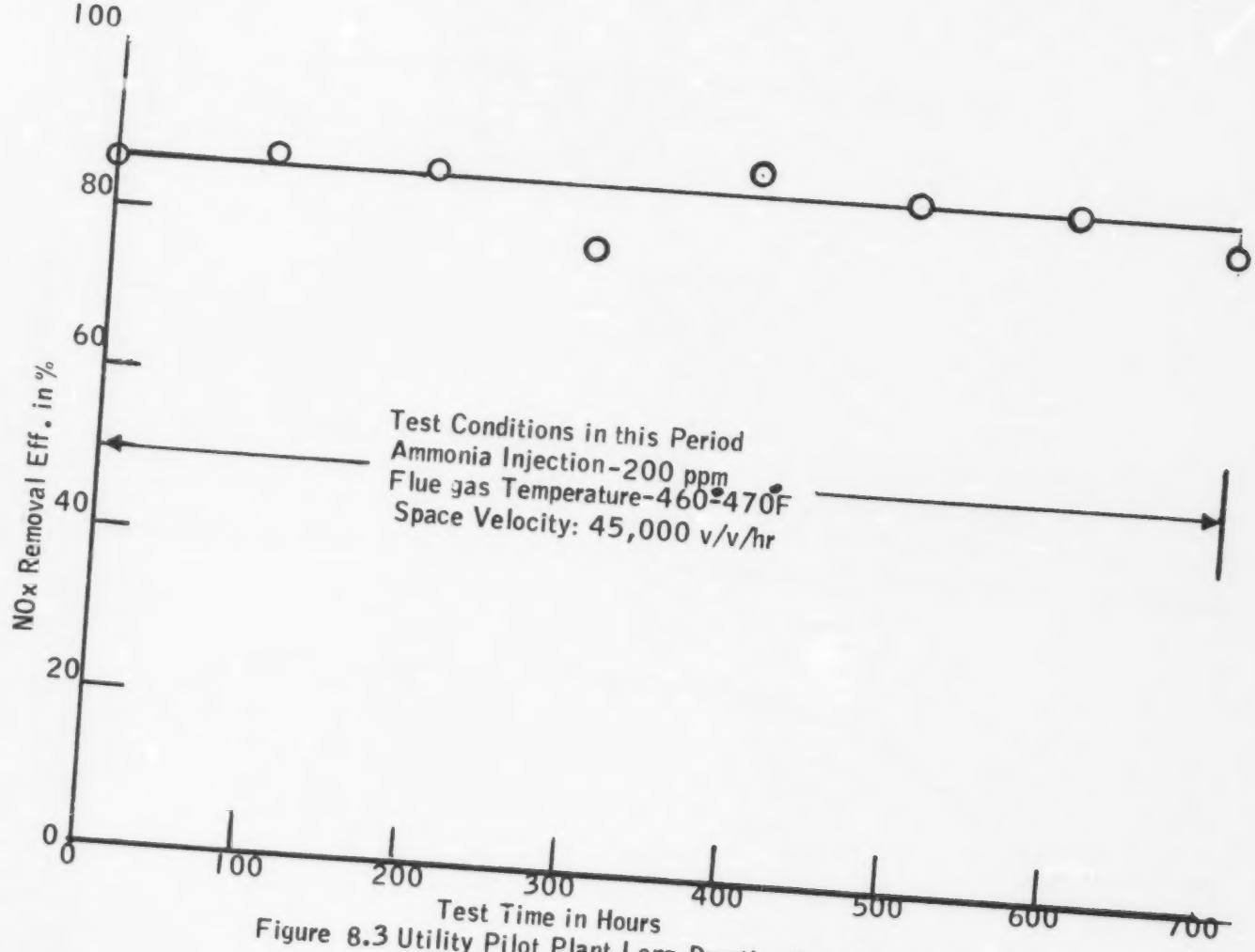


Figure 8.3 Utility Pilot Plant Long Duration Test Results

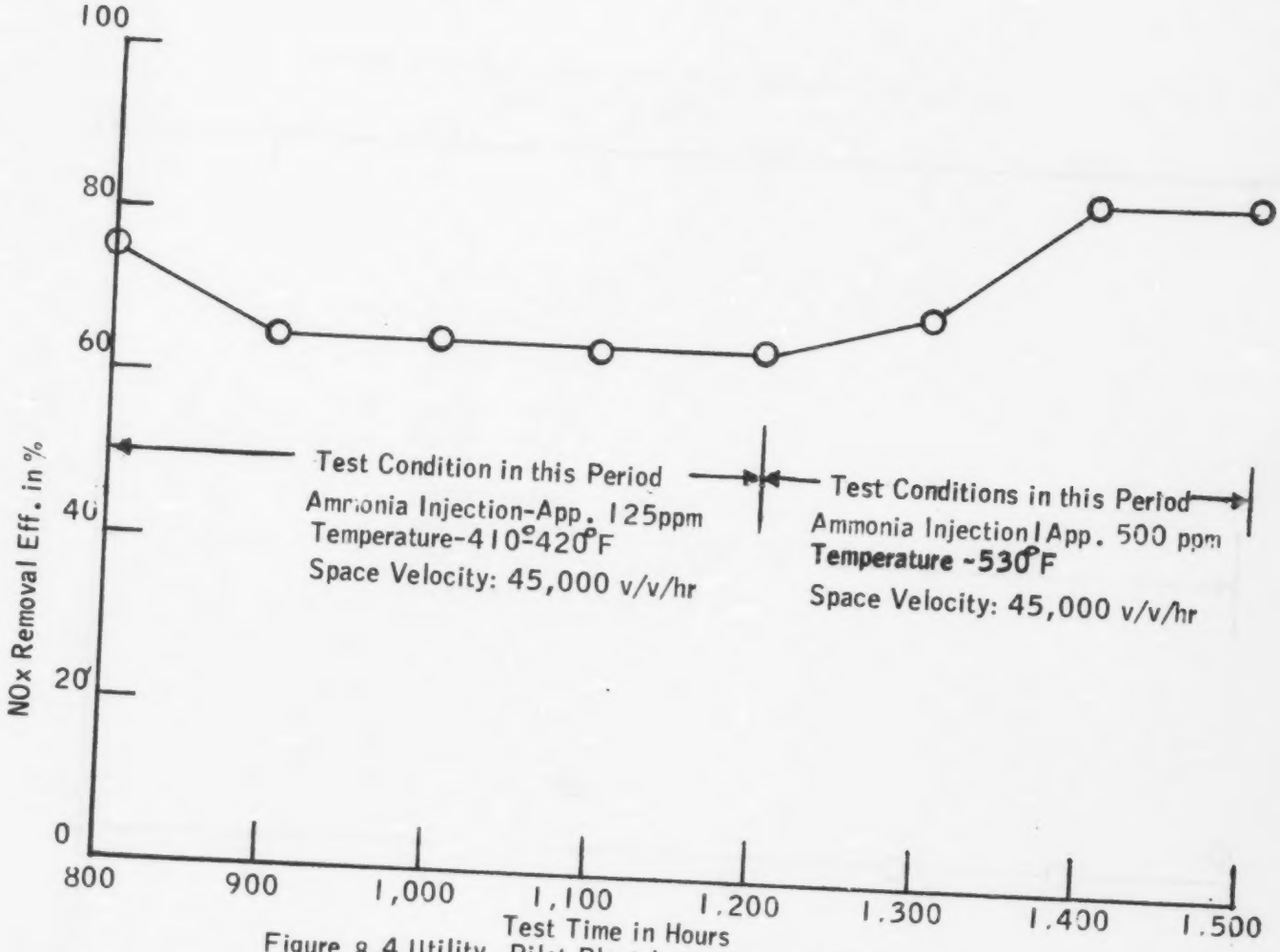


Figure 8.4 Utility Pilot Plant Long Duration Test Results

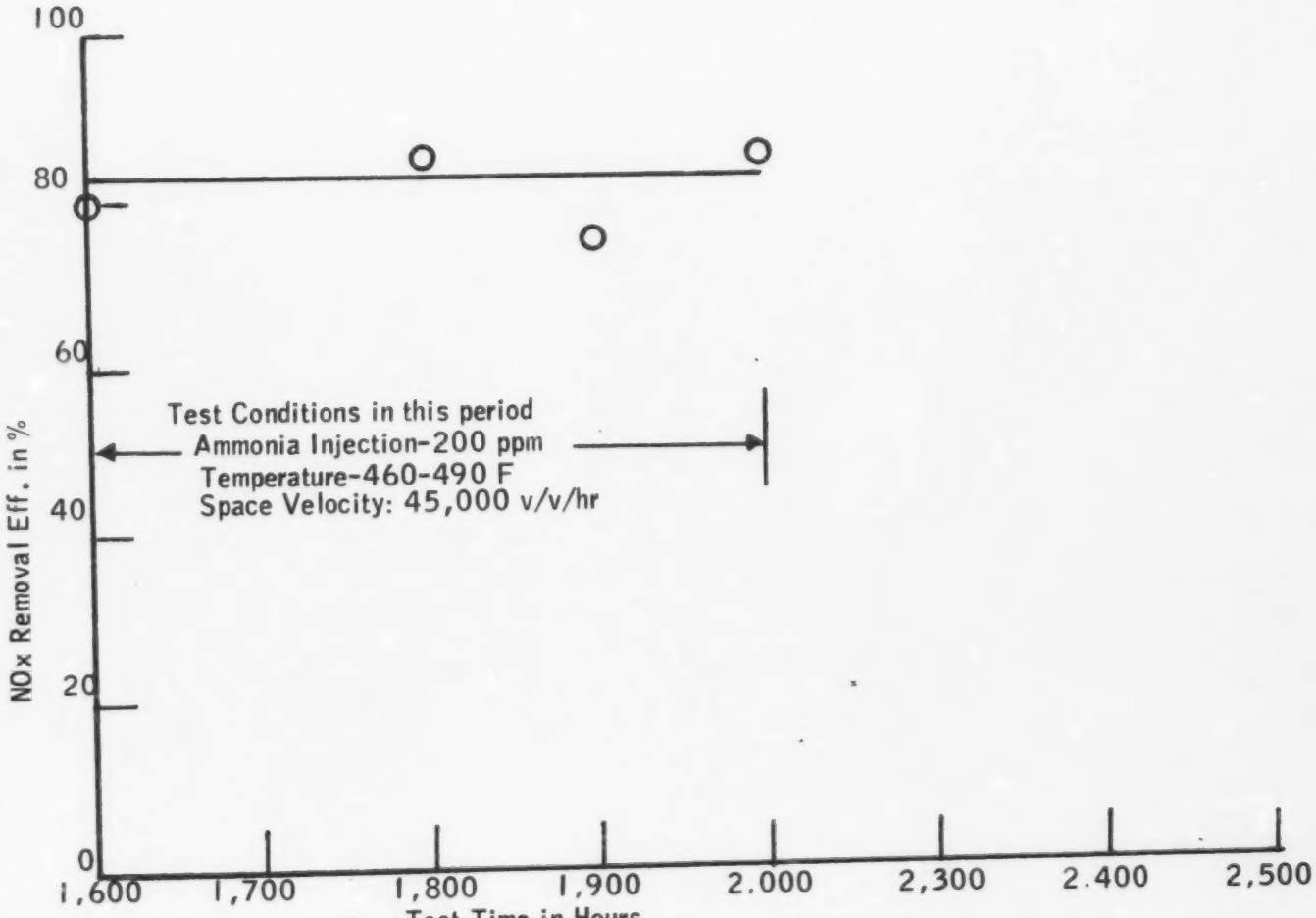


Figure 8.5 Utility Pilot Plant Long Duration Test Results

SECTION 9

OIL FIRED UTILITY PILOT PLANT TESTING

The original objective of the work described herein was to design, construct and operate a utility pilot plant NOx reduction system on gas fired flue gas only. As described previously, this pilot plant was operated on gas fired flue gas for 2000 hrs, showing 85-90% NOx removal efficiency.

Because of the success obtained with these gas fired tests, and in view of the need for a utility NOx removal process to be compatible with oil firing, the program was redirected to allow testing on oil fired flue gas. Environics' previous small-scale work indicated that a temperature of approximately 550°F was required for optimum NOx removal efficiency in the presence of SO_2, and that 80% NOx removal efficiency could be achieved. Since the utility pilot plant, as designed and built, could not provide temperatures above 500 to 525°F, an electric heater was installed in the pilot plant ductwork, immediately upstream of the air preheater gas inlet. This heater was a 75 kw Calrod duct heater, expected to increase the gas temperature by 90°F. Results of this work are described below.

After installation of the flue gas heater, the utility pilot plant was restarted, on oil firing. Due to the additional pressure drop in the system, imposed by the heater, the maximum flue gas flowrate that could be obtained was 110,000 SCFH (corresponding to a space velocity of 35000 hr^{-1}).

The utility pilot plant was operated on oil firing for approximately 412 hours, bringing the total catalyst age to 2499 hours. (During this time, the pilot plant was also operated for one 20-hour interval on gas firing.)

NOx removal efficiency ranged from a low of 19% to a high of 65%. The 19% NOx removal efficiency was observed with a gas temperature of 498°F and inlet NH_3 concentration of 150 ppm. The 65% NOx removal efficiency was observed with a gas temperature of 554°F, and inlet NH_3 concentration of 300 ppm.

There were several problems encountered during operation of the utility pilot plant under oil firing which prevented obtaining steady state parametric data on the effects of temperature and NH_3 concentration on NOx removal efficiency. These problems are described below.

First, there was a problem in obtaining and maintaining high catalyst temperatures. Although the flue gas heater did provide the expected temperature increase (100°F), gas temperature at the inlet to the pilot plant system was lower than expected and varied considerably with plant load. Catalyst temperature of 598°F could be obtained at full load (168 MW); however, at low load (40-50 MW), it was impossible to maintain temperatures above 515°F. Furthermore, because total system demand did not require high loading, it was impossible to maintain a temperature as high as 575°F for even 24 hours, to determine steady state NOx removal efficiency.

Attempts were made to increase the catalyst temperature by reducing the total gas flow rate through the pilot plant. Although reducing the gas flow rate increased the temperature across the gas heater, heat losses to atmosphere through the system seemed to increase. The net effect was that the air preheater exit temperature remained approximately the same. Temperatures were also observed to drop slightly during periods of rainy, cold or windy weather. It thus was concluded that the present thickness of insulation on the pilot plant is not quite sufficient for high temperature operation.

There also appeared to be some air flow through the air preheater, even when the air fan damper was completely closed. This was probably due to air leakage through the closed air inlet damper, induced by the gas ID fan drawing air across the rotor seals.

These fluctuations made a systematic investigation of the effect of temperature on NOx removal efficiency nearly impossible at the pilot plant.

Problems were also encountered in maintaining constant gas flow (and, therefore, constant space velocity) through the pilot plant. During high loading, the draft at the pilot plant inlet increased, and therefore pilot plant gas flowrate, and space velocity, decreased. At full load (168 MW), maximum obtainable flowrate was approximately 90000 SCFH. At low load, draft at the pilot plant inlet decreased, allowing flowrates up to 165000 SCFH. As described above, however, at low load gas temperature dropped significantly, preventing operation at optimum temperatures.

A third problem observed during operation of the pilot plant on oil firing was the accumulation of white powdery material in the exit ductwork. This accumulation was primarily observed after the pilot plant had been run overnight with NH_3 concentrations above 200 ppm. It is believed that during late evening and early morning operation, when the boiler was run at low load, the catalyst temperature dropped to the point where significant amounts of unreacted NH_3 passed through the catalyst. This NH_3 then reacted with the SO_2 in the flue gas to produce $(NH_4)\ SO_4$ or other ammonia/sulfur compounds. The temperature record indicated that, during low load operation overnight, catalyst temperature did drop, to below 500°F. Because of this problem, it was decided that, except for weekends, 24 hour operation of the pilot plant was possible to

provide catalyst aging data, but that the ammonia supply should be turned off except when Environics' operating personnel were in attendance. The LADWP subsequently analyzed a sample of the white material; their results (Table 9.1) indicate that it is primarily ammonium sulfate.

The fourth problem that was encountered during operation of the pilot plant on oil firing was fouling of the catalyst with soot particles. After 350 hours of operation on oil firing, a decrease in total gas flowrate was observed. At full load, with an inlet draft of 6" WC, the maximum flow obtainable was 50000 SCFH. After operating the pilot plant for several days under various load conditions, it was decided to remove and examine the catalyst baskets.

A basket was removed from the air preheater and the leading edge of the catalyst was found to be fouled, primarily with black soot. There were also some flakes of iron oxide, and some white particulate matter. At least one-half of the honeycomb passages appeared to be blocked. The fouling did not appear to extend into the honeycomb and, in fact, the trailing edge seemed to appear as clean as when first installed. Because the pilot plant gas and air flow are cocurrent through the air preheater, whereas in normal use they would be counter current, it was decided to rotate this basket 180° and reinstall it, thus simulating counter current flow. It was thought that this would possibly remove some of the fouling. After running the pilot plant for 20 more hours in this manner, this basket appeared to be slightly less fouled, but was still quite plugged.

Two baskets were then removed and returned to the Environics laboratory where the plugged leading edge was first swept with a small brush, and then compressed air was blown through the catalyst honeycomb from the trailing edge. This

procedure appeared to satisfactorily remove all the fouling material from the honeycomb. The remaining baskets were removed, cleaned at the utility pilot plant in a similar manner, and replaced. Immediately, after re-starting the pilot plant, a flow rate of 90000 SCFH was obtained, at full load, with an inlet draft of 6" WC. During the remainder of the catalyst operating time (100 hours), further fouling of the catalyst was not observed.

The results of the oil fired pilot plant testing described above were unexpected in several respects. By comparison with Environics' previous small scale data, much higher NOx removal efficiency was expected (80% compared with the average 50% observed in the pilot plant). The difficulty in obtaining high and constant gas temperatures, due to low load demand, was also unexpected. Results of the pilot plant work indicated that the optimum temperature might be above 550°F, and such temperatures could not be readily obtained. The degree of soot fouling encountered was also unexpected, as were the frequent maintenance shutdowns of the boiler.

It was therefore decided to redirect the program to laboratory pilot plant testing, on oil fired furnace flue gas. This would permit more extensive determination of the effect of temperature than possible in the pilot plant. Results of this work are described in Section 10.

Table 9.1

Department of Water & Power
of the City of Los Angeles
Power System
ma

Report No. C-1658
Page 1 of 2

TESTING LABORATORIES

SPECIAL TEST & INVESTIGATION REPORT

DEPOSIT

Environics' Pilot Plant

Valley Steam Plant, Unit No. 4

A sample of a deposit, removed on February 3, 1975 from the exhaust duct of Environics' Pilot Plant at Valley Steam Plant, Unit No. 4, was submitted by Mr. Charles Sunwoo for chemical analysis.

The purpose of the pilot plant is to determine the effectiveness of the reduction of oxides of nitrogen by the addition of ammonia vapor to the stack gases in the presence of a platinum catalyst.

The unit had been gas fired for several months without accumulation of any noticeable deposit. After fuel oil firing for only a few days a large amount of the submitted material had collected.

Most of the deposit consisted of flat flakes, with diameters ranging from 1/16 to 3/8 inches. Many were of a rusty brown or white to yellow color on both sides, others were rusty on one side and white or yellow on the other side. It also contained magnetic particles.

The weight loss of the sample was determined at 105°C. The loss at this temperature is primarily water of hydration. The sample was then heated to 815°C and the weight loss determined. The loss at the higher temperature accounted for over half the weight of the sample.

When heated to the higher temperature the sample first

Date Completed: 3/7/75 Test Made by: AEC
Authorization: Report by: AEC
Job Card No. 24320-T-12 Checked by:
Copies to: W. W. Pepper (3)
E. L. Morrison (1) Approved: s/W. E. Greninger

Table 9.1 (continued)

emitted dense white fumes. They were strongly basic and it was confirmed that ammonia was being liberated. When heating was continued the reaction of the fumes changed from basis to acidic, indicating that sulfur trioxide was being liberated. These two gases are decomposition products of ammonium sulfate.

The ashed residue of the sample was analyzed and was found to consist primarily of iron oxide. Small amounts of constituents associated with fuel oil were also present. However, the large amount of iron, compared to the other elements present, indicated that most of the deposit is the result of corrosion.

By convention the metallic constituents are reported as their respective oxides, but are also present as other compounds. More than half of the sample was soluble in water and about 99% was soluble in dilute hydrochloric acid. The sulfates were determined on the acid soluble portion of the deposit.

Data

Determinations on sample as received

	Percent by weight
Weight loss at 105 C	3.89
Weight loss at 815 C	53.16
Iron as Fe_2O_3	38.94
Silica SiO_2	0.95
Zinc as ZnO	0.46
Vanadium as V_2O_5	0.30
Calcium as CaO	0.25
Magnesium as MgO	0.23
Nickel as NiO	0.22
Sodium as Na_2O	0.19
Aluminum as Al_2O_3	0.14
Copper as CuO	0.02
Water solubles	54.33
Acid solubles	98.86
Sulfates as SO_3	38.00

SECTION 10

OIL FIRED LABORATORY PILOT PLANT TESTING

The laboratory pilot plant tests were conducted using the 1.2×10^6 BTU/hr boiler, used for the NH_3 mixing tests and catalyst selection tests described in Section 5. The boiler was modified to burn #2 fuel oil, and to heat and control the flue gas temperature over a range from 500 to 640°F.

Combustion gases are exhausted from the boiler stack by an ID fan and pass through the heater section, mixing section, and catalytic converter. All ductwork downstream of the fan is 6" square. The heater section contains 6 finned "Calrod" heaters, rated at 2 kw each. The heaters are wired in pairs, two of which are switched manually, and the third by a temperature controller. The flue gas temperature at the catalyst bed inlet can be controlled to any desired temperature between 500°F, with none of the heaters on, and 640°F, with all heaters on. Ammonia and NO are injected into the center of the boiler stack, so that they may mix with the flue gas during passage through the fan and downstream ductwork. (Injection of bottled NO is required to increase the flue gas NO concentration from 75 ppm produced by the burner, to 125 ppm expected in this utility pilot plant boiler.)

Total flue gas flowrate is approximately 12,500 SCFH, and is measured by an orifice plate at the duct exit. Since the catalyst bed dimensions are 6" x 6" x 12" (catalyst volume is 0.250 ft^3), the catalyst operates at a space velocity of 50,000 v/v/hr.

Inlet and exit gas samples are drawn from the center of the duct upstream and downstream from the catalyst bed. Samples are analyzed for NOx, O_2, and SO_2 by a Beckman 951

Chemiluminescence NOx Analyzer, Beckman 715 Process O_2 Monitor, and EnviroMetrics NS200 SO_2/NOx Analyzer (for SO_2).

The first tests conducted with the laboratory pilot plant used a batch of platinum catalyst, 6" square x 12" thick. The catalyst was ordered from Matthey Bishop, Inc., suppliers of the pilot plant catalyst, and was to be produced in a manner identical to that of the utility pilot plant catalyst.

First, the effect of inlet NH_3 concentration on NOx removal efficiency was determined, at three different temperatures. Results of these tests are shown in Figure 10.1. As can be seen, NOx removal efficiency is, in general, approximately the same at 540 and 580°F, and significantly less at 610°F, at the NH_3 concentration tested. Furthermore, NOx removal efficiency continues to increase with increasing NH_3 concentration up to 1000 ppm, although the increment decreased beyond 550 ppm.

Next, the effect of continued exposure to oil fired flue gas on NOx removal efficiency was explored. Using the data of Figure 10.1, the conditions initially chosen were 500 ppm NH_3 and 540°F.

At 540°F, NOx removal efficiency dropped from approximately 70% to 50% after about 8 hours. At this point, temperature was increased to 580°F. Removal efficiency increased to 70%, but, after approximately 32 hours, dropped to 60%. At that point, temperature was again increased, to 610°F. Removal efficiency remained constant for 20 hours, at 65%, but then decreased to 40% after 60 hours of operation at 610°F (total age, 140 hours). The NOx removal efficiency at 580°F was checked and found to be 20%, compared to 60% when the catalyst age was 60 hours. Activity at a higher temperature (640°F) was also tested, and found to be 30%. At that time, oil fired testing of Catalyst #1 was stopped.

The catalyst was then removed from the converter, and, in

an attempt to regenerate it, was heated at 700°F, overnight, in an electric kiln. Following this procedure, the catalyst was run on gas firing to determine its NOx removal efficiency as a function of temperature. After 72 hours, the catalyst reached the steady state performance. The NOx removal efficiency peaked at 85% at a temperature of 525°F. This is quite different from the performance of the utility pilot plant catalyst where peak efficiency of 90% at 475° to 500°F has been observed. Tests on this catalyst were discontinued and the catalyst manufacturer was contacted to determine whether this batch was indeed identical to that supplied for the pilot plant. It was then discovered that the manufacturing process was slightly different from that used originally. Although the resulting catalyst was considered identical, insofar as precious metal loading and substrate characteristics are concerned, this may account for the observed discrepancies.

The next step in the laboratory pilot plant work was to remove some catalyst from the utility pilot plant, for testing in the laboratory pilot plant. Two catalyst baskets were removed from the air preheater and the catalyst was cut to provide two 6" square x 12" thick beds for the laboratory pilot plant.

Catalyst from basket #3 of the utility pilot plant air preheater was tested first, in the oil fired laboratory pilot plant, with the results shown by the circles in Figure 10.2. Figure 10.2 shows the effect of inlet NH_3 concentration on NOx removal efficiency, and Figure 10.3 the effect of temperature on NOx removal efficiency. The effect of NH_3 concentration was first determined at 560°F. Other operating conditions, constant for all tests, were: space velocity, 50,000 hr^{-1} and inlet NOx concentration: 100 ppm. The results were somewhat surprising, in that NOx removal efficiency continued to increase as NH_3 concentration was increased to 1000 ppm. It was

decided to determine the effect of temperature using an NH_3 inlet concentration of 300 ppm. Results of this work are shown by the circles in Figure 10.3. At this level, peak NOx removal efficiency of 45% occurred at 600-620°F. It appeared that the catalyst's performance was rapidly decreasing, compared with the average of 50% observed for several hundred hours during oil fired operation of the utility pilot plant. Furthermore, the performance peak occurred at significantly higher temperatures than observed previously.

The catalyst from basket #2 was then loaded into the laboratory pilot plant, and tested in a manner similar to the first, with the results shown by the squares in Figure 10.2. The performance of basket #2 catalyst was similar to, but somewhat lower than, that of basket #3.

On the theory that operation in the rotating air preheater in the pilot plant (alternate exposure to flue gas and warm air) provided a continuous regeneration effect, delaying the deactivation of the catalyst, basket #3 catalyst was regenerated. The regeneration procedure consisted of placing the catalyst in an electric kiln held at 500°F for 66 hours. After this procedure, the catalyst was replaced in the reactor and tested, with the results shown in Figure 10.4.

The first tests performed were to determine the effect of temperature on NOx removal efficiency, at 300 ppm inlet NH_3 concentration, under oil firing. Results are shown by the circles in Figure 10.4 It can be seen that the catalyst performance was significantly improved by the regeneration procedure. Peak NOx removal efficiency increased to 58% (from 45%) and the temperature peak shifted to 580 to 600°F (from 600-620°F). After operating for 14 hours at 580°F, the catalyst performance had dropped slightly, from 54% to 48%. At this point, it was decided to switch the boiler to gas firing

in order to determine whether the catalyst performance would return to that observed on gas firing before exposure to oil fired flue gas. Therefore, the gas burner was installed in place of the oil burner on the lab pilot boiler and tests were made to determine the effects of temperature on NOx removal efficiency under gas firing, with other conditions identical to those described above. Results are shown by the squares in Figure 10.4.

It appears that the effects of exposure to oil firing are reversible; i.e., after a switch to gas firing, the catalyst gradually reverts to its performance prior to oil firing.

The data from the oil fired laboratory pilot tests on the utility pilot plant catalyst is, in several respects, contrary to what has been observed in Environics' previous small-scale (2" dia.) tests. In these previous experiments, the catalyst, identical to that used in the current work, exhibited 80% NOx removal efficiency at 575°F, and 300 ppm NH_3, after 450 hours exposure to furnace flue gas containing 300 ppm SO_2. The major difference between the previous and current work, other than scale, is that the early work was conducted with flue gas from a gas fired furnace, to which bottled SO_2 was injected. It is possible that the flue gas in the current oil fired boiler tests contains compounds other than SO_2 (such as heavy metal compounds) which could "poison" the catalyst.

Although no definite conclusions can be drawn, it does suggest that the platinum catalyst may be susceptible to poisoning by some component of oil fired flue gas, other than SO_2. Using a platinum catalyst in actual oil fired flue gas, NOx removal efficiency of 50% can be obtained in a regenerative system such as an air preheater wheel, at temperatures between 550 and 600°F. The activity of the catalyst does appear to be restored when the boiler is returned to gas firing.

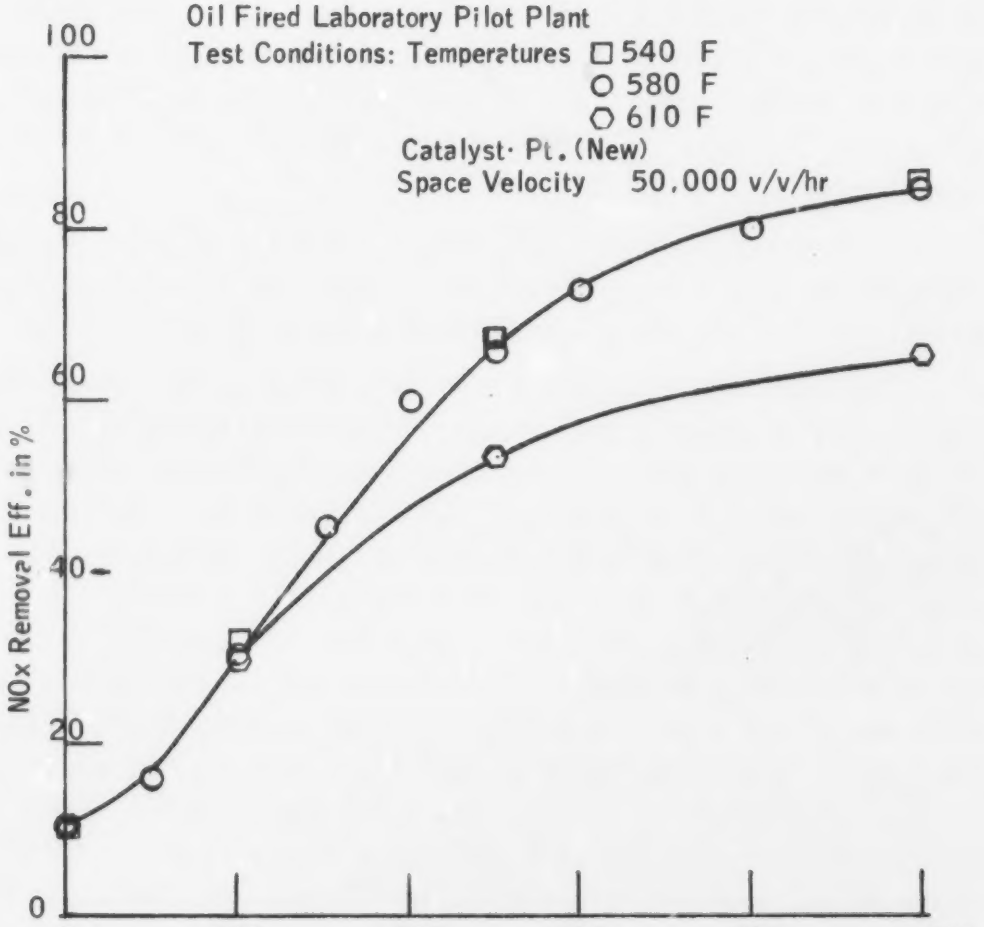


Figure10.1 NOx Removal Eff. Vs Inlet Ammonia Conc.
New Catalyst (Pt)

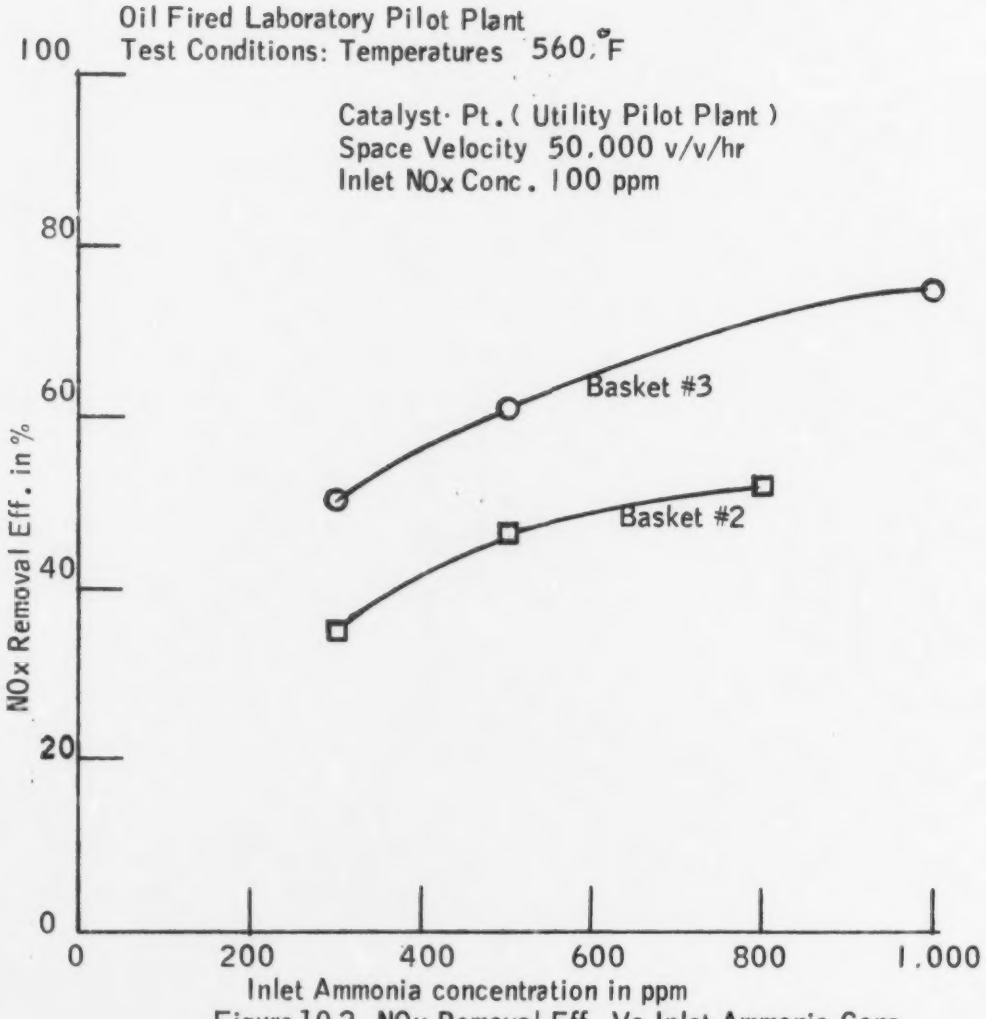


Figure 10.2 NOx Removal Eff. Vs Inlet Ammonia Conc.
(Utility Pilot Plant Catalyst)

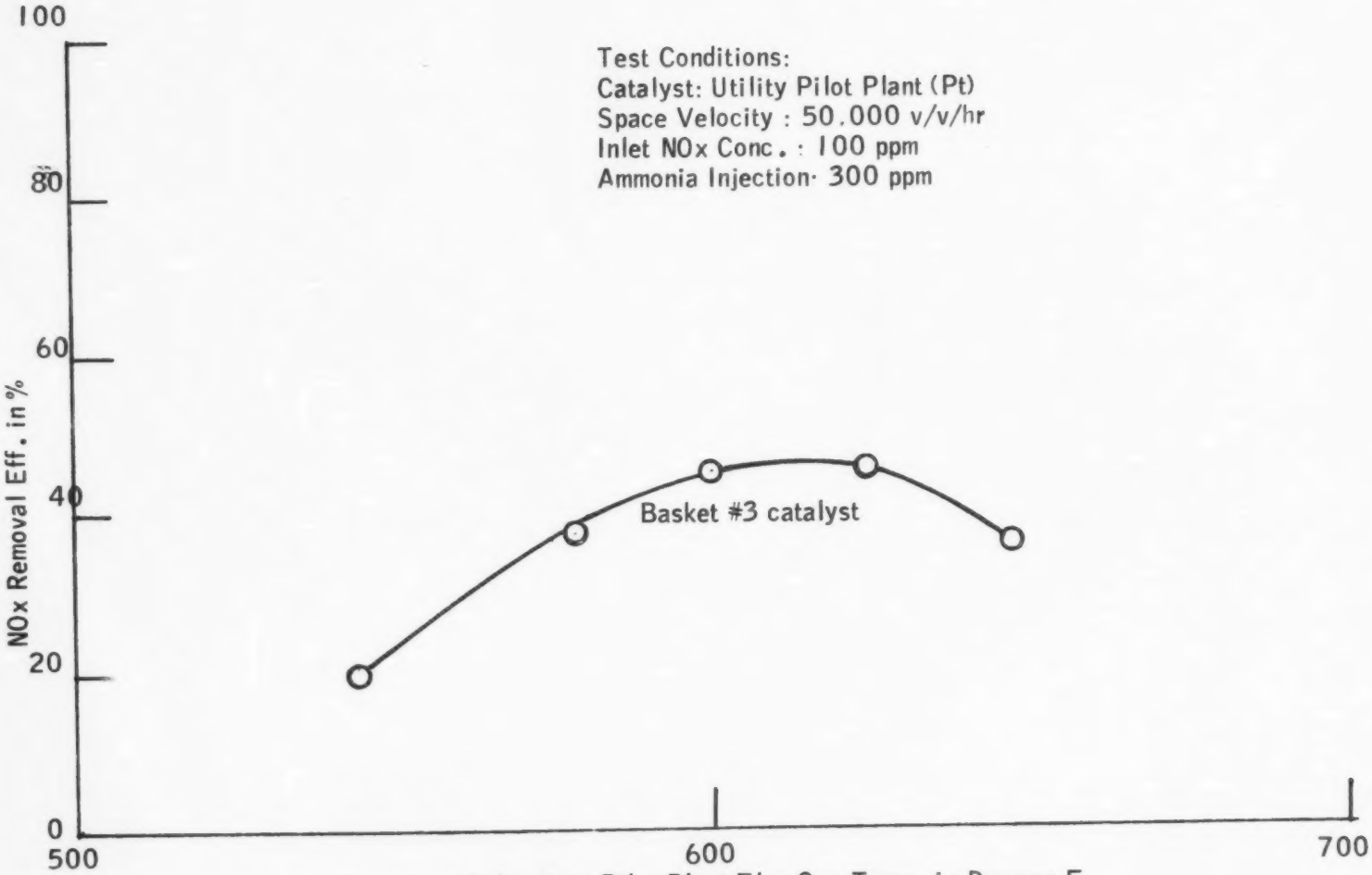


Figure10.3 NOx Removal Eff. Vs Temp.
Laboratory Pilot Plant

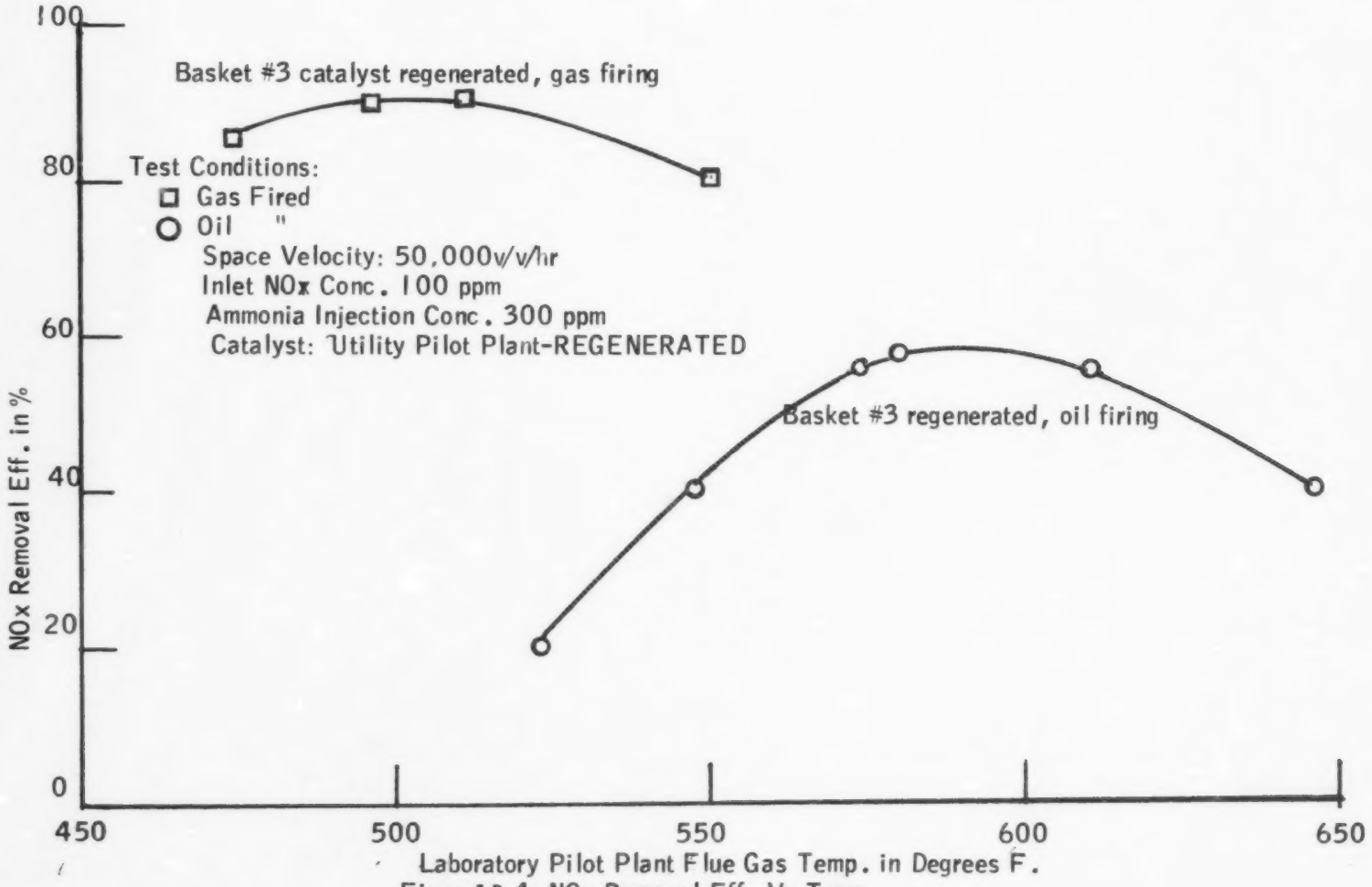


Figure10.4 NOx Removal Eff. Vs Temp.
Laboratory Pilot Plant

SECTION 11

COST ESTIMATE-480 MW PLANT

It was desired to prepare a cost estimate for a full scale catalytic NOx reduction system, as part of the pilot plant program. A preliminary cost estimate for a full scale gas fired boiler NOx reduction system is therefore presented below, based upon the following assumptions.

1. The system will be installed on a 480 MW plant, flowing 50×10^6 SCFH of flue gas.
2. The concentration of NOx in the flue gas will be 125 ppm.
3. The catalyst cost will be $1,000 per cubic foot.
4. Ammonia will cost 18 cents per pound, including tankage.
5. The catalyst bed will be designed for operation at 50000 hr^{-1} space velocity.
6. The catalyst will be installed in the middle baskets of the existing air preheater.
7. The catalyst bed will be replaced at 5 year intervals; 25% of the replacement catalyst cost can be offset by reclamation of the precious metal in the spent bed.
8. The cost of capital is 10%.

Cost of the system can be broken down as follows:

1. Capital Costs
   - A. Catalyst bed
   - B. Catalyst containment system
   - C. Instrumentation and control system

2. Operational Costs

A. Ammonia consumption

B. Maintenance expense

The estimated capital expenditure schedule is shown in Table 11.1. These estimates are discussed separately, below.

1. Catalyst Bed Cost

The required catalyst bed volume (in the flue gas stream) is 1000 cubic feet. Because of its location in the air preheater, only 45% of the catalyst is actually in the gas stream at any given time. Therefore, 2222 ft^3 of catalyst are actually required. A catalyst cost of $1000 per ft^3 has been quoted for purchase of such quantities. The catalyst cost is therefore $2,222,000. The best projection that can be made at this time is that the catalyst bed would have to be replaced no more often than every five years. It is probable that 25% of the replacement catalyst cost could be defrayed by recovery of the precious metal from the spent catalyst. Replacement beds would thus cost $1,666,500 each. The present value of the original catalyst bed and five replacement beds is thus $4,668,000 over the 30 year life of the plant.

2. Catalyst Containment

The catalyst containment system will hold the catalyst modules in place in the air preheater wheel, to provide protection from mechanical shock, prevent gas from bypassing the catalyst, and allow access to sections of the catalyst bed for cleaning and/or replacement. Cost of this system is estimated to be $500,000.

3. Instrumentation and Control System Cost

It is estimated that the instrumentation and control system will initially cost approximately $20,000 and have a life of 10 years. The present value of instrumentation is then $30,700 over the 30 year life of the plant.

4. Operation Cost

Ammonia will be consumed at the maximum rate of 455 lb/hr. Ammonia cost is estimated to be 18 cents per pound, or $82 per hour. Thus, for a 480 MW steam plant, ammonia cost would amount to 0.017 cents per kilowatt-hour. Additional operational cost will come from maintenance of the system. It is estimated that they will require less than one man full time, or $20,000 per year (0.0005 cents per kilowatt-hour).

A cost summary for the catalytic reduction system is presented in Table 11.2. Standard present value methods were used. The capital cost per KW of plant capacity is $10.83. The operating cost is 0.0176¢ per KW hour.

CAPITAL OUTLAY SCHEDULE

	Years						
	0	5	10	15	20	25	30
Catalyst	$2,222,000	1,666,500	1,666,500	1,666,500	1,666,500	1,666,500	(555,500)*
Catalyst Containment	$ 500,000						
Instrumentation	$ 20,000		20,000		20,000		

*denotes salvage value of last catalyst bid.

Table 11.1

TABLE 11.2
COST SUMMARY
FOR 480 MW STEAM PLANT

1. CAPITAL COST

Present Value of Catalyst Cost	$4,668,000
Present Value of Catalyst Containment Cost	500,000
Present Value of Instrumentation Cost	30,680
Total Present Value Cost	$5,198,680
Total Present Value Cost, per KW of Plant Capacity	$10.83

2. OPERATING COST

Total Maintenance Cost (per year)	$ 20,000
Maintenance Cost (per KW hour)	0.0005¢
Ammonia Cost (per KW hour)	0.0171¢
Total Operating Cost per KW hour	0.0176¢

APPENDIX A

CONVERSION FACTORS

Environmental Protection Agency policy is to express all measurements in agency documents in metric units. When implementing this practice will result in undue cost or lack of clarity, conversion factors are provided for the non-metric units used in a report. Generally, this report uses British units of measure. For conversion to the metric system, use the following conversions:

To convert from British units	To Metric units	Multiply by
SCFH	NCMH (Normal Cubic Meter Per Hr)	2.832×10^{-2}
v/v/hr	v/v/hr	1
ppm	ppm	1
Inch	Centimeter	2.540
Feet	Meter	0.3048
square inches	square centimeters	6.452
Inches of water	mm of Mercury	1.868
RPM	RPM	1
%	%	1
MW	MW	1
$°F = \frac{9}{5}\ °C + 32$		
Btu per hour	Kilo-calories per hour	0.2520